Light Water Reactor Sustainability Program

Techno-Economic Analysis of Synthetic Fuels Pathways Integrated with Light Water Reactors

Techno-Economic Analysis of Synthetic Fuels Pathways Integrated with Light Water Reactors

L. Todd Knighton, Daniel S. Wendt (INL)

Lesley Snowden-Swan, Jeromy Jenks, Shuyun Li, Steven Phillips, and Jalal Askander (PNNL)

September 2020

Page intentionally left blank

EXECUTIVE SUMMARY

Synthetic fuels (synfuels) and chemicals (synchems) are produced by synthesis from chemical building blocks rather than by conventional petroleum refining. Synthesis gas or syngas (carbon monoxide and hydrogen) is a common intermediate building block in the production of synfuels and synchems. Syngas can be produced by many processes, including biomass or fossil fuel gasification and by co-electrolysis. In co-electrolysis, CO_2 is reacted with water to produce syngas. The CO_2 can be sourced from processes that would otherwise eject the CO_2 to the atmosphere, such as ethanol plants, including dozens of large plants in the United States that convert corn into ethanol that is being blended with the national gasoline, or fossil fuel processes, such as steam methane reforming or natural gas combined cycle (NGCC) power plants. CO_2 is also emitted from biofuels gasification plants. Conversion of CO_2, which would have otherwise been released to the atmosphere, to synfuels using nuclear energy is a potential avenue for adding value to existing light water reactor (LWR) facilities, while producing transportation fuels that are compatible with conventional fuels produced via petroleum refining. The cost of CO_2 separation depends on the purity of the source. Valorization of CO_2 is a critical complementary component of carbon capture and utilization (CCU) and an alternative to carbon capture and sequestration (CCS).

The purpose of this work is to identify, model, perform techno-economic analysis, and compare two possible synfuel production routes utilizing CO_2 as the feedstock. Heat from an LWR nuclear plant is integrated wherever possible to positively affect the economics of the LWR by converting power to fuels during times of low grid electricity demand. Process and economic modeling for a conceptual synfuel production plant co-located (or in near proximity) with an LWR is presented, including the cost of CO_2 captured from an ethanol plant, compressed, and transported to the LWR hybrid plant, co-electrolysis of the CO_2 with water in a solid oxide electrolyzing cell (SOEC) system to produce syngas, and thermo-catalytic conversion of the syngas to transportation fuel. The hybrid LWR/synfuels plant is assumed to be located within 50–150 miles of an ethanol plant (e.g., located in the midwest region of the United States). Performance and n^{th}-plant economics for the co-electrolysis-based processes are evaluated and compared with biomass-gasification-based technology for the synfuel routes considered. Sensitivity analysis around the price of CO_2 and electricity, two of the major cost drivers, is presented for each case. Consideration of a carbon credit is also included in the sensitivity analysis. The primary results and conclusions of the analysis are the following:

- For a plant producing 3,195 barrels per day (BPD) hydrocarbon synfuels via a methanol intermediate with LWR electricity and steam usage of 326 MWe and 133 MWt respectively:

  - The modeled minimum fuel selling price (MFSP) of diesel (91%) and gasoline (9%) blendstock with conservative assumptions is \$4.45/gallon for the base case using a \$33.3/tonne CO_2 cost and \$30/MWh electricity price. That is compared with the biomass gasification route to syngas with an MFSP of \$3.28/gallon. Note that co-electrolysis has a much larger maximum scale of production that can be reached compared to the availability of land competing with food production as in the case of biomass gasification. Also, the scale of the analysis is only about 1/3 of the available energy from a typical LWR but was chosen so that a direct comparison with a biorefinery could be made.

  - There are innovative cryogenic carbon capture (CCC) processes that claim to produce CO_2 for a cost as low as \$20 to \$60/tonne CO_2 (SES 2020), which could also have significant impact on the viability of an LWR/synfuels plant using methanol as intermediate. Further, the refrigerant used in the CCC process could be produced using LWR energy. The synergies of the LWR with the CCC process and techno-economic modeling of the CCC process will be explored in detail in future studies.

  - Sensitivity analysis (Figure ES-1) shows that with optimal CO_2 and electricity prices and inclusion of carbon credits through incentives or mandates this process could be more cost

competitive with petroleum fuels, especially post Covid-19 when oil prices recover somewhat from the current historic lows. With a hypothetical carbon tax of $100/tonne CO_2, the MFSP is reduced to ~$3.75/gallon. A renewable fuel standard (RFS) credit would further aid in competitiveness of fuels produced via this route. Some states already offer credits for clean fuels, including California and New York. These credits are qualified under the U.S. Environmental Policy Act and are applied to select fuels with the assignment of Renewable Identification Numbers (RINs). Clean fuels credits in California have ranged upwards from $0.5 to $2.5 per gallon of gas equivalent.

- Sensitivity analysis varying plant scale for the co-electrolysis with methanol-to-olefins (MTO) fuel process was conducted (Figure ES-2). At a scale of half the base case (326 MWe; 133 MWt), production cost increases by 9%. At a scale 10 times larger than the base case, production cost is reduced to about $3.8/gal. Scaleup of the plant up to the entire electrical output of a general 1-GWe LWR of fuel production would result in about 40 cents/gal cost savings. Note that a scaling factor of 1 is assumed for the SOEC stack; therefore, no benefit is gained for this portion of the capital cost.
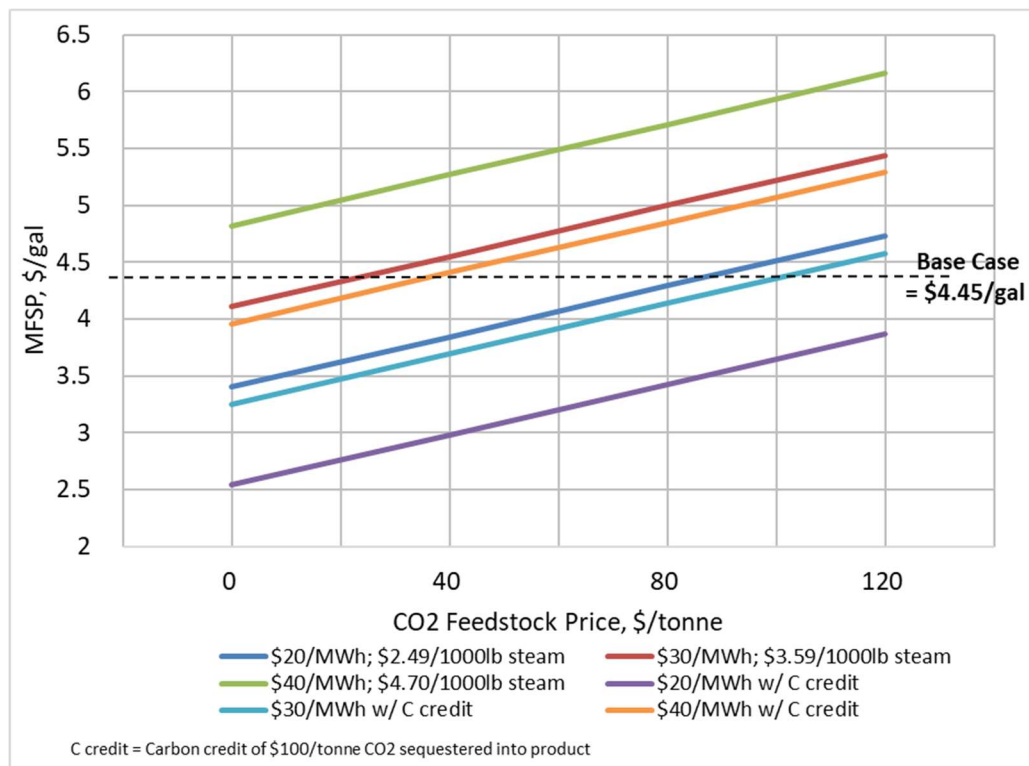

Figure ES-1. Minimum fuel selling price sensitivity for fuels via a co-electrolysis derived syngas to methanol to fuels route.
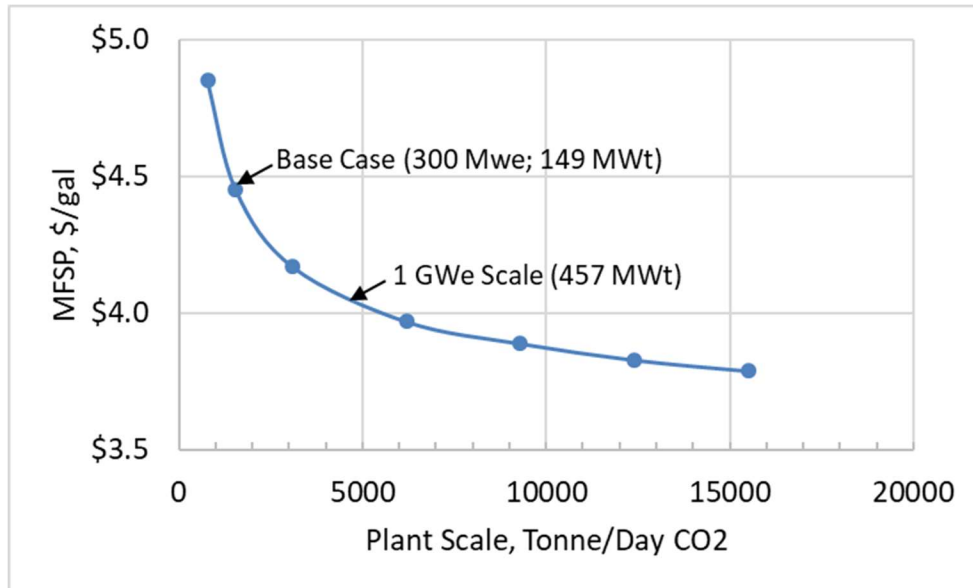
Figure ES-2. Sensitivity of MFSP to plant scale for the co-electrolysis and MTO fuel process.

- The ethanol (EtOH) pathway is considered to be a low-to-mid technology readiness level (TRL) level technology, as the mixed alcohol conversion process to produce ethanol has only been tested at pilot scale. For a plant producing 2,870 BPD hydrocarbon synfuels via an ethanol intermediate with LWR electricity and steam usage of 468 MWe and 66 MWt, respectively:

  - The modeled MFSP is $6.13/gallon using a $33.3/tonne CO_2 cost (including compression and transportation from an ethanol plant to the LWR/synfuels plant) and $30/MWh electricity price. This option has a higher cost compared to the methanol intermediate route primarily because the thermochemical syngas conversion to mixed alcohols process is only about half as carbon efficient at making ethanol as the syngas-to-methanol process.

  - Compared with biomass gasification route, the MFSP of the co-electrolysis case with ethanol intermediate is about 52% higher.

  - Sensitivity analysis (Figure ES-2) indicates it will be somewhat more challenging to make this pathway cost competitive with petroleum fuels even considering optimal feedstock cost and carbon credits.

  - By using a syngas to ethanol process with a higher carbon efficiency (e.g., fermentation) economics for this pathway could potentially be improved.
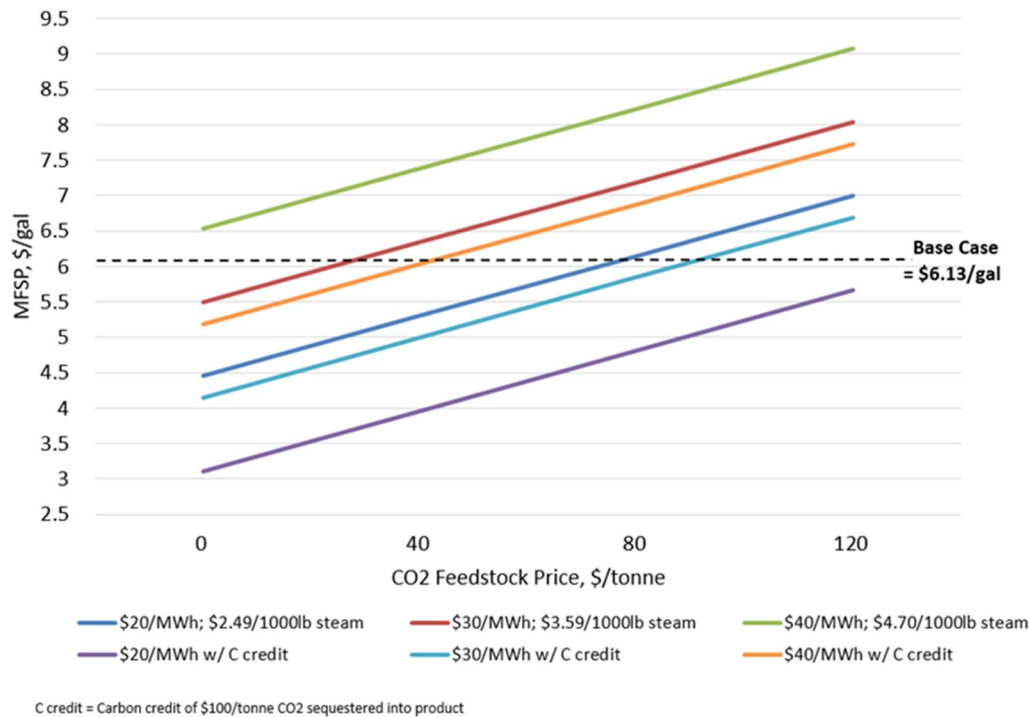
Figure ES-3. Minimum fuel selling price sensitivity for fuels via a co-electrolysis derived syngas to ethanol-to-fuels route.

The results of this study justify further pursuit of synfuels via the methanol-to-fuels as an alternative market for LWR energy use. Co-electrolysis could take all of the energy provided by a single reactor or two reactors to produce the syngas that is converted to methanol. The synfuels could be competitive in price with petroleum fuels when credits for CO_2 emissions reductions reach about $100/tonne CO_2 or when the price of petroleum fuels rises above the current historic lows. The combination of plant scale-up matching the energy produced by an average nuclear power plant, plus clean energy credits would make synthetic fuels produced by co-electrolysis using LWR energy competitive with petroleum-derived fuels. Together, biomass gasification and nuclear-derived synfuel could feasibly replace a significant volume of U.S. transportation fuels. The nation currently burns 12 million barrels of gasoline and diesel each day. Biomass gasification and co-electrolysis together can feasibly replace over 25% of the petroleum fuels.

Future studies should take into account opportunity sources of CO_2, their purity, and location, financial investment terms and options, and clean energy credits. In addition, synergies between nuclear power plants and the biomass gasification synfuels route should be considered, including biomass feedstock drying and torrefaction, and CO_2 by-product from biomass gasification.

Page intentionally left blank

CONTENTS

FIGURES

TABLES

Page intentionally left blank

ACRONYMS

ACCE	Aspen Capital Cost Estimator
ASR	area-specific resistance
BEA	Battelle Energy Alliance
BETO	Bioenergy Technology Office
BMI	Battelle Memorial Institute
BPD	barrels per day
CAPEX	capital expenses
CCC	Cryogenic carbon capture
CCS	carbon capture and sequestration
CCU	carbon capture utilization
CE	chemical engineering
DME	dimethyl ether
DOE	Department of Energy
EPA	Environmental Protection Agency
ER	electrochemical reduction
EtOH	ethanol
FA	formic acid
FCI	fixed capital investment
FT	Fischer-Tropsch
GHG	greenhouse gas
GREET	Greenhouse Gases, Regulated Emissions, and Energy Use in Transportation model
INL	Idaho National Laboratory
LCFS	low-carbon fuel standard
LWR	light water reactor
LWRS	Light Water Reactor Sustainability program
MAS	mix alcohol synthesis
MeOH	methanol
MFSP	minimum fuel selling price
MTG	methanol can also be converted to hydrocarbons via the methanol-to-gasoline
MTO	methanol-to-olefins
MTP	methanol to propylene
NG	Natural gas
NGCC	Natural gas combined cycle

NPP	Nuclear power plants
OPEX	operating expenses
PNNL	Pacific Northwest National Laboratory
REC	Renewable energy credits
RFS	renewable fuel standard
RIN	renewable identification number
RSTOIC	stoichiometric reactor in AspenTech software
RWGS	reverse water-gas-shift
SES	Sustainabile Energy Solutions
SOEC	solid oxide electrolyzer cell
SOFC	solid oxide fuel cell
TEA	techno-economic analysis
TRL	technology readiness level
UOP	Honeywell UOP, formally known as Universal Oil Products
YSZ	yttria stabilized zirconium
ZEC	Zero emissions credits

1.   INTRODUCTION

Nuclear energy is increasingly being recognized as a valuable low-carbon, low-emissions energy source that can help achieve clean energy targets being set by states, commissions, and utilities in the United States. Currently, nuclear power provides about one-fifth of the country's electricity. Nuclear power plants (NPPs) further provide the grid with all-weather season-long baseload capacity that is important to grid reliability and resiliency. Light water reactor (LWR) NPPs in the United States, like other sources of electricity generation, are facing increasing cost pressure on the electricity grid due to historically low-priced natural gas (NG) and the rapid expansion of solar and wind energy. Solar and wind energy provide spikes on the grid during periods of high production, but there will be a continued opportunity for baseload generators, such as NPPs, to provide electricity to the grid when solar and wind energy installations are producing little output. During times of grid overgeneration the NPP energy can be diverted to create other value-added chemical and fuels. Therefore, the U.S. Department of Energy (DOE) Light Water Reactor Sustainability (LWRS) Program is addressing flexible plant operations that can diversify the revenue of NPPs.

Previous reports have evaluated opportunities to couple LWRs with hydrogen production (Knighton 2020a, Knighton 2020b, Hu 2019, Frick 2019). This report analyzes several synthetic fuel (synfuel) production pathways that could be coupled with LWRs to provide alternative options to utility companies for using nuclear energy to create value added products during periods of overgeneration of electricity to the grid. A conceptual integrated plant would consist of a hybrid LWR delivering power and heat to a synthetic fuel and/or chemical facility. The synfuels plant would employ co-electrolysis to convert CO_2 and water into syngas (synthesis gas, a mixture of H_2 and CO). The CO_2 ideally would come from a source that is in close proximity to the LWR and one in which the CO_2 is currently being released to the atmosphere, to take advantage of possible clean energy credits. Sources such as an ethanol plant release CO_2 in high concentration which makes the CO_2 separation and utilization more cost effective. In this report an ethanol plant located between 50 and 100 miles from the LWR is assumed to be the CO_2 source. CO_2 sources in close proximity to LWRs, such as a natural gas combined cycle (NGCC) plant, and state-of-the-art carbon capture technology will be evaluated in future work. The syngas would then be converted to synthetic fuels via the most economical processes. Choices for the conversion of syngas to synfuels include Fischer-Tropsch (FT), methanol-to-gas (MTO), ethanol-to-fuels, as well as other possible options. This report focuses on the evaluation of two possible syngas to fuels pathways: (1) methanol-to-olefins (MTO) and (2) ethanol-to-fuels. It is recognized that there are other possible pathways to fuels as well as pathways to valuable synthetic chemicals from syngas that could be analyzed in future work.

Co-electrolysis is assumed to take place with a solid oxide electrolyzing cell (SOEC) to take advantage of its high efficiency as compared with a standard polymer electrolyte membrane (PEM) electrolyzer. Although higher value synchems offer compelling investment potential, the focus of the techno-economic analysis (TEA) detailed herein is on production of synfuel blendstocks compatible with existing liquid transportation fuels. The methanol (MeOH) and ethanol (EtOH) routes were selected because methanol and ethanol are common intermediates that can either serve as base chemicals or be converted to hydrocarbon fuels; therefore, the analysis is flexible to be used for future synchems analyses. In addition, Pacific Northwest National Laboratory (PNNL) has extensive experience in syngas-based fuels modeling and TEA so existing biomass-based PNNL models for the methanol and ethanol-based conversion routes could be leveraged for this work.

The plant scale chosen is commensurate with the typical scale used for a cellulosic biorefinery (2000 dry ton/day biomass feed) to provide a consistent comparison with renewable fuel from biomass. This is equivalent to a syngas flow of 141 ton/day H_2 and 973 ton/day CO for the methanol-to-fuels-based pathway and 162 ton/day H_2 and 1,909 ton/day CO for the ethanol-to-fuels-based pathway (see Sections 5 and 6). Detailed process models and TEA were developed for synfuel pathways incorporating the use of the LWR heat and electricity as an energy source for the conversion process. Equipment CAPEX and

OPEX are detailed in the economic modeling, including reactor and other unit operation costs, and feedstock and product valuations. Sensitivity analysis around key process and economic assumptions is also presented.

This report begins with a high-level overview of the various possible routes from CO_2 to synfuels and chemicals in Section 2 and a description of the two pathways selected for detailed modeling and analysis. Section 3 outlines the general approach and underlying assumptions for the TEA. Section 4 details the models generated for predicting syngas generation from an SOEC plant and the costing of the SOEC stack. Section 5 and 6 presents the process design and TEA results of the fuel pathways via the syngas-to-methanol and syngas-to-mixed alcohols (primarily ethanol) routes, respectively. Section 7 discusses the current and possible future structure of national and state level carbon credit systems and how these systems may improve the economics of synthetic fuels produced integrated with clean nuclear energy. Section 8 summarizes the results and conclusions for the two fuel routes evaluated.

2.   ROUTES FOR CO_2 TO FUELS OR CHEMICALS

There is a myriad of possible ways to make fuels or chemicals from CO_2. This section is intended to give a high-level overview of the some of the possible technology options and selection of the two pathways that were analyzed in detail. Figure 1 provides a flowchart of possible pathways for producing CO_2-derived fuels integrated with an LWR. Steam and power from the LWR are provided for CO/H_2 (syngas) production in addition to any demand required by the chemical plant. Syngas can be produced from CO_2 via co-electrolysis and the reverse water-gas-shift (RWGS) reaction. Co-electrolysis is preferred over having separate electrolyzers for CO_2 and water as it has found to be more efficient (Fu et al. 2010). Syngas is used to produce a wide range of fuels and chemicals, including but not limited to synthetic NG, dimethyl ether, methanol, ethanol, and hydrocarbon fuel blendstocks. Ethanol and methanol can also be produced directly with co-electrolysis. However, it is thermodynamically unfavorable compared to making syngas, as the number of electrons needing fixed is 6 for methanol and 12 for ethanol, versus 2 for CO (Verma et al. 2019). Table 1 lists the numbers of electrons required for CO and other chemicals.
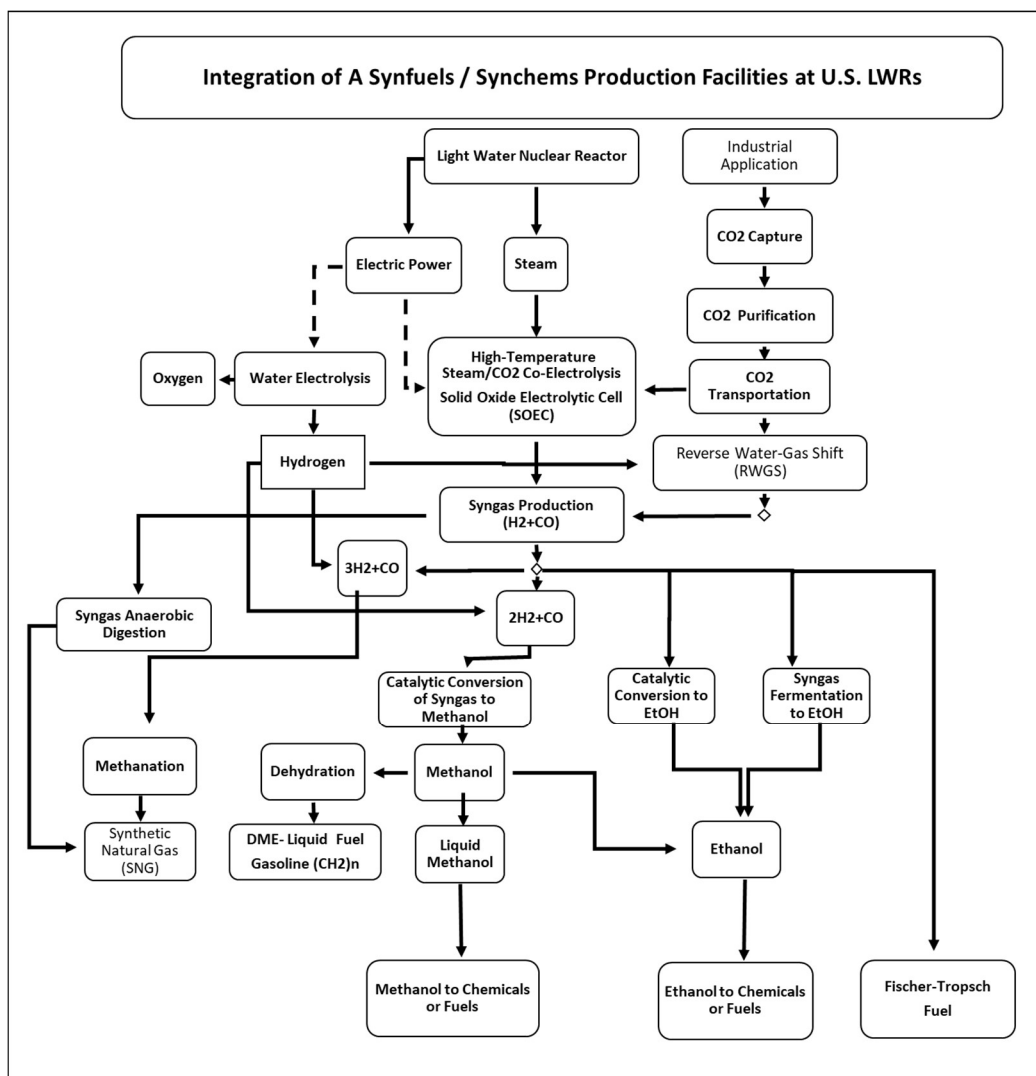
Figure 1. Overview of various synfuel/synchem pathways integrated with an LWR.

Table 1. Number of electrons transferred for converting CO_2 to syngas and chemicals.

Compound	Electrons Transferred
Syngas	2 e-
Formate	2 e-
Methanol	6 e-
Methane	8 e-
Ethanol	12 e-
Octanol	48 e-

2.1.1 Fuels

Syngas can be converted to oxygenate fuel, such as ethanol or methanol, or converted further into hydrocarbon fuels. Ethanol can be produced from syngas through biological or thermochemical means, as

shown in Figure 2. Fermentation has been indicated by many researchers as an efficient and cost-effective method for conversion of syngas and several companies have now made the process commercial. LanzaTech now commercially produces ethanol via syngas fermentation using its proprietary Clostridial biocatalyst. Direct thermo-catalytic conversion of syngas to ethanol is also possible using several different catalysts, as shown in Figure 3.
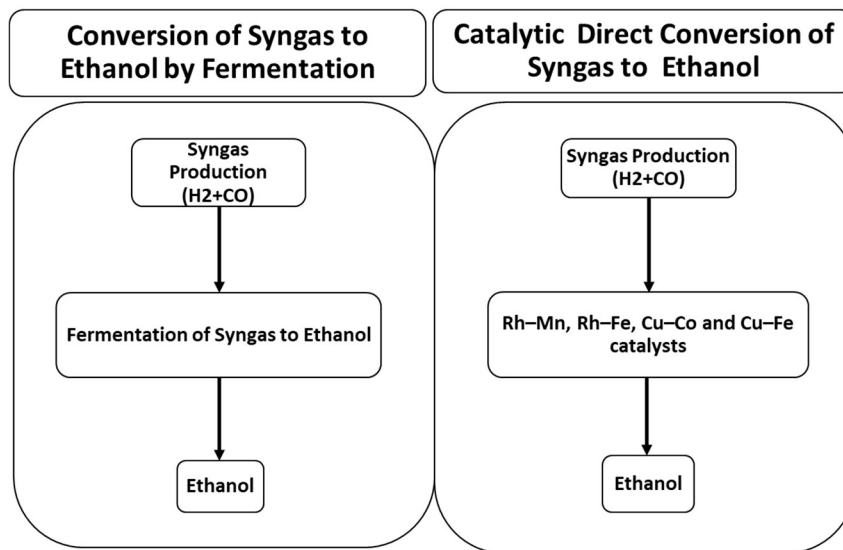
Figure 2. Fermentation and catalysis of syngas to ethanol.

Figure 3 details three alternative pathways for syngas conversion to ethanol that are in earlier stages of development than direct conversion shown in Figure 2. Figure 3a shows a direct conversion of syngas to ethanol through dimethyl ether (DME) as a key intermediate, the catalyst $ZnAl_2O_4$ / H-MOR $ZnAl_2O_4$ produces ethanol with a selectivity of 52% (Zhou et al. 2018). There are some other fuel alternatives, such as DME, which is a clean-burning, non-toxic, potentially renewable fuel that can be produced from methanol. The path of methanol-to-ethanol production is not necessarily the cheapest. Some other tracks with inexpensive catalysts may be worth evaluating like methyl acetate. The two-step conversion of methanol to ethanol via the methyl-acetate process is at the stage of pre-commercialization. Figure 3b details industrial methanol carbonylation to produce acetic acid that is performed either over the Ir-based (Cativa process) or Rh-based (Monsanto process) catalysts (Lu et al. 2016). Figure 3c shows the methanol conversion process through dimethyl oxalate. In this process ethylene glycol and ethanol can be produced, making it a versatile method for producing essential chemicals (Yue, Ma, & Gong, 2014).
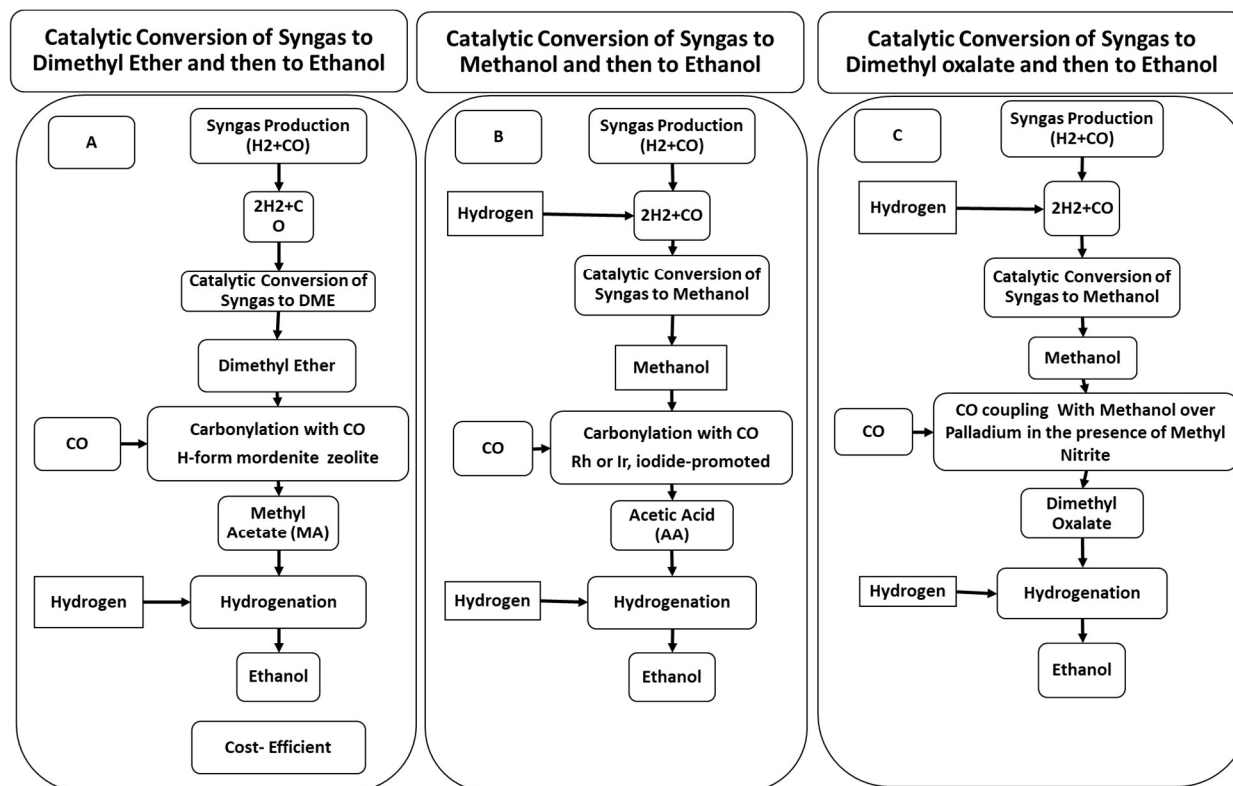
Figure 3. Syngas conversion to ethanol via catalysis.

Methanol can also be converted to hydrocarbons via the methanol-to-gasoline (MTG) process or the MTO process, both of which were originally introduced by the Mobil Oil Corporation. MTG is carried out over a HZSM-5 catalyst with high selectivity and little side-products, producing a hydrocarbon mixture of narrow compositional range (Gogate 2019). The MTO process was developed by essentially controlling process conditions to interrupt the MTG methanol to hydrocarbons reaction and has been commercialized by Honeywell UOP (UOP) and others.

Syngas can also be converted directly into hydrocarbon fuel using the established FT process, as shown in Figure 4. The FT process is the oldest coal-to-liquids technology, invented in the 1920s and used by the Germans during World War II to provide needed liquid hydrocarbon fuels (NETL 2020). Several FT-based commercial plants are operating today, including Sasol's Sasolburg coal-to-liquids plant (South Africa). Sasol started developing designs for a gas-to-liquids plant in Lake Charles, Louisiana, but the project was cancelled in 2017 due to the collapse of oil prices which decreased the differential between the NG feedstock and the value of the fuel products (Griggs 2017).
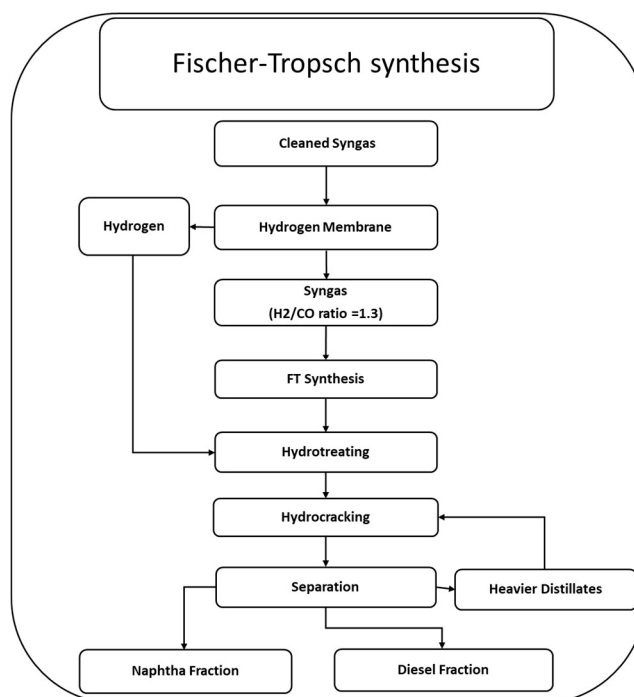
Figure 4. Syngas to hydrocarbon fuel via Fischer-Tropsch synthesis.

2.1.2   Chemicals

Many pathways exist for converting CO_2 to chemicals via thermochemical or electrochemical processes followed by catalysis. CO_2 reduction reactions can yield various valuable multi-carbon compounds including ethylene, acrylic acid, propylene, and C1 chemicals and polymers (Alper & Yuksel Orhan, 2017). As of 2017, 130 Mt of CO_2 is used annually to generate urea, salicylic acid, polycarbonates, and cyclic carbonates (Plasseraud 2010). Figure 5 shows two pathways for converting methanol to propylene or ethylene. The MTO reaction detailed in Figure 5b is one of the most critical reactions in C1 chemistry, which provides a chance for producing basic petrochemicals such as ethylene and propylene (Eng 1998). The methanol-to-propylene (MTP) process shown in Figure 5a produces propylene from methanol (Koempel & Liebner, 2007). Significant differences exist between the MTO and MTP processes in regard to reactor design and productivity (Barger 2002). MTO uses a fluidized-bed reactor, where heat can be removed quickly, and catalysts can be easily regenerated. MTP uses a fixed bed reactor where heat removal is problematic but overcome by using multiple catalyst beds. Fixed bed reactors are easier to scale up compared to fluidized beds, are cheaper and have better product selectivity. However, MTO can use crude methanol whereas MTP must use pure methanol, thus adding to the overall cost for MTP (Jasper & El-Halwagi, 2016). As is evident, the chemical derivatives of ethylene and propylene are numerous and have a variety of industrial applications. Polymerization processes are not detailed here.

6
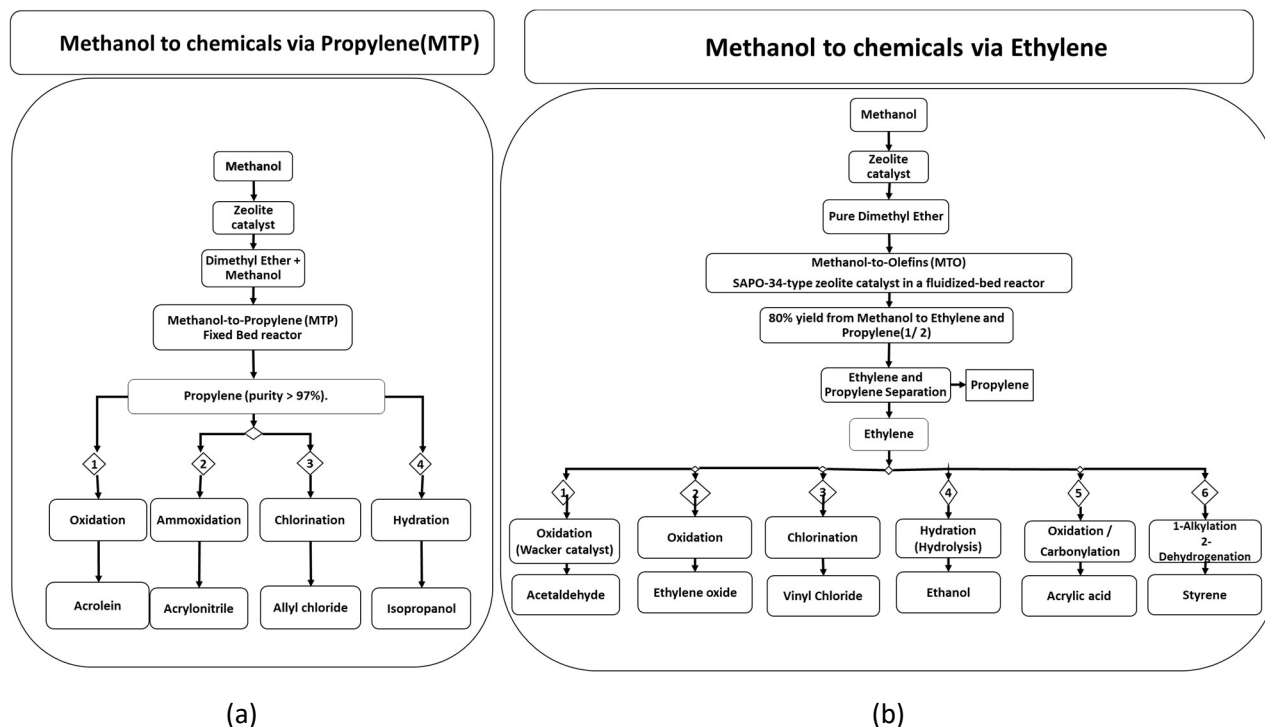
Figure 5. Methanol conversion to chemicals

Ethanol is an essential source of many chemical compounds, including para-xylene. Ethylene is the intermediate compound in the process of producing para-xylene, either directly (Figure 5b), or indirectly (Figure 6a). In a direct path, ethylene is used in more than one step of a complex process to produce para-xylene (Zhang, Qian, Kong, & Wei, 2015). In the indirect method, DMF is the primary compound for the production of para-xylene, and ethylene plays a role in forming double bonds (Lyons, Guironnet, Findlater, & Brookhart, 2012).
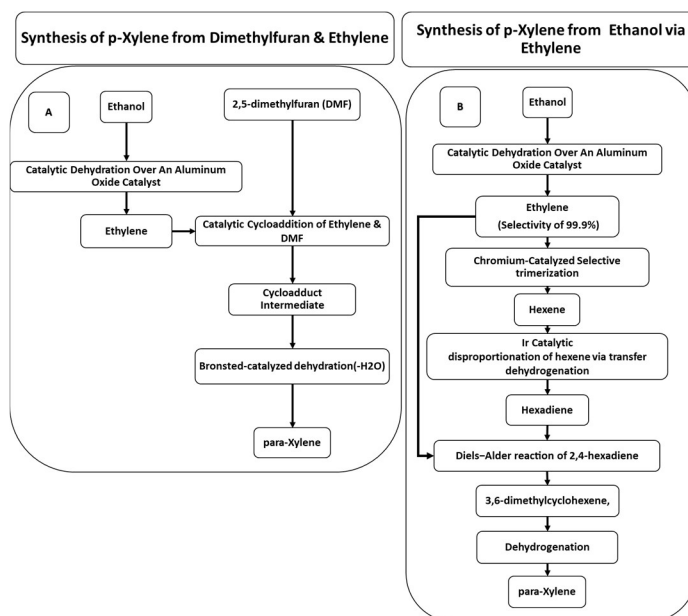
Figure 6. Ethanol conversion to para-xylene.

As a final note on chemical conversion, formic acid (FA) is a critical commodity used in agricultural, pharmaceutical, food, textile, and other chemical markets. The global demand for FA is expected to be 820,000 metric tons in 2021 (Sesto 2016). FA can be produced via electrochemical reduction (ER) or homogenous catalysis of CO_2 and H_2. Processing requirements range from 25 to 400 MJ/kg of FA produced (Rumayor, Dominguez-Ramos, & Irabien, 2018). This would require less than one and up to three LWRs dedicated solely to the production of FA to meet global demand.

2.1.3 Selected Hydrocarbon Fuel Routes for TEA

As the focus of this study is transportation fuels, conversion pathways converting CO_2 to fungible hydrocarbon fuel were selected for detailed modeling and TEA. Production of syngas via co-electrolysis or RWGS result in the same overall chemical equation (Eq. 1); therefore, similar energy consumption. Co-electrolysis via SOEC was selected as Idaho National Laboratory (INL) and PNNL have ongoing R&D in this area. As discussed, the possible range of syngas-to-fuel pathways is very extensive. From this extensive list, the possible range of technologies was down-selected to an ethanol- and methanol-based pathway, as both are versatile chemicals that can be further converted into fuels, and a wide range of chemicals and products (Dagle et al. 2020). As such, future studies building on this work may enable evaluation of chemical products or co-products, which can significantly improve the process economics (Dagle et al. 2020). In addition, existing process models previously developed by PNNL for biomass gasification and conversion to synfuels via the methanol and ethanol-based pathways could be leveraged and adapted.

$$4 H_2O + 2 CO_2 + 12 e^- \rightarrow 4 H_2 + CO + 3 O_2 \tag{1}$$

The pathways selected for detailed analysis are shown in Figure 7. The methanol pathway is based on established technology for methanol generation from syngas and production of olefins from methanol using the UOP's commercialized MTO process (Bipin V. Vora, D. et al. 1998). Olefin oligomerization and hydrogenation technology are based on a PNNL-patented process (Lilga, et al. 2016). The methanol pathway is considered to be at a high technology readiness level (TRL), as the syngas-to-methanol technology (using coal or NG) has been in use for decades and there are several industrial installations of the MTO process based on coal gasification, one being in China and one in Belgium (Gogate 2019). The ethanol-based model is based on thermochemical syngas-to-mixed alcohol conversion technology (Stevens et al 1989; Dutta et al 2011) and the ethanol-to-butene process is currently being developed by PNNL (Dagle et al. 2020a). Olefin oligomerization and hydrogenation steps are based on the PNNL patented process, consistent with the methanol process model. The ethanol pathway is considered to be a low-to-mid TRL level technology, as the mixed alcohol conversion process to produce ethanol has only been tested at pilot scale (Summers et al 2019), and the ethanol-to-butene process is still in research stages.
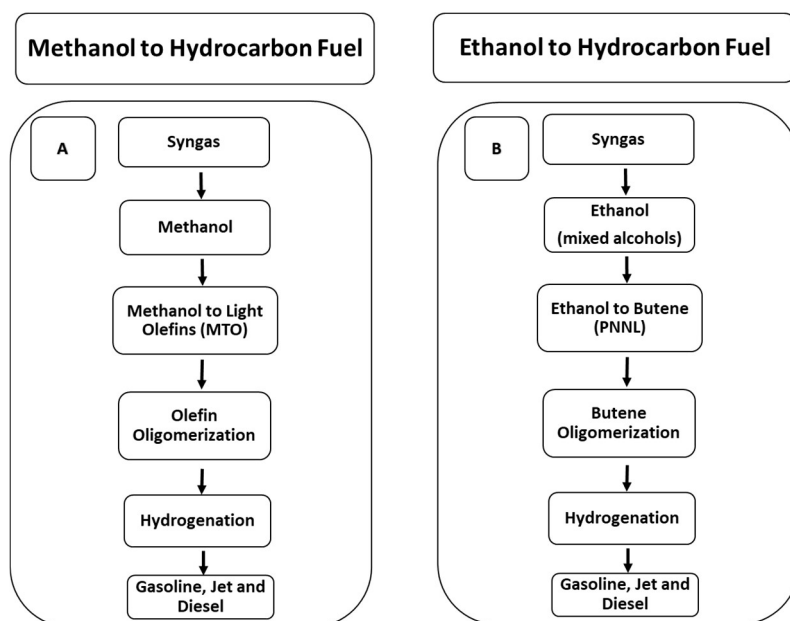
Figure 7. Selected syngas intermediate chemicals and pathways for fuel production through (a) methanol and (b) ethanol.

3. TECHNO-ECONOMIC ANALYSIS METHODOLOGY

The approach to developing conversion process techno-economics is similar to that employed in previous analyses conducted for the DOE's Bioenergy Technologies Office (BETO) (Dutta et al. 2015; Jones et al. 2013, 2014; Tan et al. 2015). Process flow diagrams and models are developed based on experimental research by PNNL, INL, and others, along with information from the literature and commercial vendors for mature and similar technologies. To assure consistency across all conversion pathways, BETO developed a set of economic assumptions that are used for all bioenergy TEAs (DOE 2016), which are also adapted for this work. An important aspect of these assumptions is that they reflect an "nth-plant" design, as described below.

3.1 Definition of Nth Plant

A standard reference basis common to the conceptual design reports, known as the "nth" plant design, is used. These assumptions do not account for additional costs that would normally be incurred for a first-of-a-kind plant, including special financing, equipment redundancies, large contingencies, and longer startup times necessary for the first few plants. For nth-plant designs, it is assumed that the costs reflect a future time when the technology is mature, and several plants have already been built and are operating. The specific assumptions are shown in Table 2. Note that tax incentives and other credits that may be applicable (e.g., credits under the Renewable Fuel Standard or cellulosic biofuels bonus depreciation) but are excluded from the analysis to represent plant economics independent of any government subsidies.

Table 2. Nth-plant assumptions.

Assumption Description	Assumed Value
Internal rate of return (IRR)	10%
Plant financing debt/equity	60%/40% of total capital investment (TCI)
Plant life	30 years
Income tax rate	21%

Interest rate for debt financing	8.0% annually
Term for debt financing	10 years
Working capital cost	5.0% of fixed capital investment (excluding land)
Depreciation schedule	7-years MACRS[a] schedule
Construction period	3 years (8% 1st yr, 60% 2nd yr, 32% 3rd yr)
Plant salvage value	No value
Startup time	6 months
Revenue and costs during startup	Revenue = 50% of normal Variable costs = 75% of normal Fixed costs = 100% of normal
On-stream factor	90% (7,920 operating hours per year)

(a) Modified accelerated cost recovery system

3.2 General Cost Estimation Basis

All costs in this report are on a 2019 constant dollar basis. This is the current reference year that BETO uses to facilitate comparison of various conversion technologies (DOE 2016). Capital costs are estimated from a variety of resources. The heat and material balances generated by the simulation software (ASPEN-Plus [AspenTech 2013]; CHEMCAD v.7) are used to size the major pieces of equipment. Aspen Capital Cost Estimator (ACCE), information from published literature and vendors quotes are used to cost individual pieces of equipment. The original cost reflects the year of the cost quote or estimate, and the scale of the equipment. All capital costs are adjusted to an annualized 2019 basis using the Chemical Engineering magazine's published indices:

$$\text{Cost in 2019 \$ = equipment cost in quote year} \times \left(\frac{2019 \text{ index} = 541.7}{\text{quote cost year index}} \right) \quad (2)$$

The scale is adjusted to match the appropriate scaling term (heat exchanger area for example) by using the following expression:

$$\text{Scaled equipment cost = cost at original scale} \times \left(\frac{\text{scale up capacity}}{\text{original capacity}} \right)^n \quad (3)$$

where n is the scale factor, typically, 0.6 to 0.7.

After equipment is scaled and adjusted to the common cost year, factors are applied to calculate the total capital investment. Individual installation factors calculated by ACCE are multiplied by equipment costs, unless installed costs are already available from vendors. The total direct cost is the sum of all the installed equipment costs, plus the costs for buildings, additional piping, and site development. Indirect costs are estimated as 60% of the total installed costs. Factors for the calculation of these additional direct and indirect costs are listed in Table 3. The sum of the direct and indirect costs is the fixed capital investment (FCI). The total capital investment is the fixed capital plus working capital and land costs.

Table 3. Cost factors for direct and indirect costs.

Direct Costs	
	% of Total Installed Cost (TIC)
Buildings	4.0%

Site development	10.0%
Additional piping	4.5%
Total Direct Costs (TDC)	18.5%
Indirect Costs	
	% of TDC
Prorated expenses	10%
Home office and construction fees	20%
Field expenses	10%
Project contingency	10%
Startup and permits	10%
Total Indirect Costs	60%
Working Capital	5% of FCI
Land	HTL: 6 acres @ $15,000/acre Upgrading: 6% of Total Purchased Equipment Cost

Operating costs are estimated by using the results from the ASPEN-Plus heat and material balances and applying raw material and utility prices (given in individual pathway TEAs in following sections). Labor requirements and rates for the modeled synfuels plant are consistent with past TEAs performed for biomass-gasification-based fuel plants and are listed in Table 4. Note that labor needs associated with running the front-end SOEC portion of the plant as compared to a gasifier may be lower; therefore, these costs are likely conservative.

Table 4. Labor costs for modeled synfuels plants.

Fixed Operating Costs	$/Year	No. workers	Base yr $/hr	$/y Total in 2011$	$/y per Worker in 2019$	$/y Total in 2019$	$/hr in 2019$
Plant Manager	161,362	1	70.67	161,362	170,761	170,761	82.10
Plant Engineer	76,839	1	33.65	76,839	81,315	81,315	39.09
Maintenance Supr	62,569	1	27.40	62,569	66,214	66,214	31.83
Lab Manager	61,471	1	26.92	61,471	65,052	65,052	31.27
Shift Supervisor	52,690	5	23.08	263,450	55,759	278,795	26.81
Lab Technician	43,908	3	19.23	131,724	46,466	139,397	22.34
Maintenance Tech	43,908	16	19.23	702,528	46,466	743,449	22.34
Shift Operators	**43,908**	**27**	**23.08**	**1,185,516**	**46,466**	**1,254,570**	**22.34**
Yard Employees	30,736	12	13.46	368,832	32,526	390,316	15.64
Clerks & Secretaries	39,517	3	17.31	118,551	41,819	125,456	20.11

With the capital and operating costs, the MFSP is determined using a discounted cash flow rate of return analysis. The MFSP is the plant gate selling price of the fuel product that makes the net present value of the project equal to zero given the financial factors assumed (see Table 2).

4. CO-ELECTROLYSIS OF CO₂ AND WATER TO SYNGAS

SOECs offer a unique method for converting CO_2 and steam into syngas. Co-electrolysis is preferred over separately electrolyzing steam and CO_2 because of reduced cell resistance (area-specific resistance) and lower conversion of CO to C (Stoots 2010). Figure 8Figure 8 details the process inclusive of the SOEC stack, heat exchangers, compressor, pump, and separators. In this process CO_2 and water enter the stack in vapor phase. Power supplied to the stack yield syngas and oxygen. Oxygen flows from the cathode to the anode and is removed by a sweep gas, typically air. Excess CO_2 and steam are separated from the stack output and reintroduced to the stack inlet.
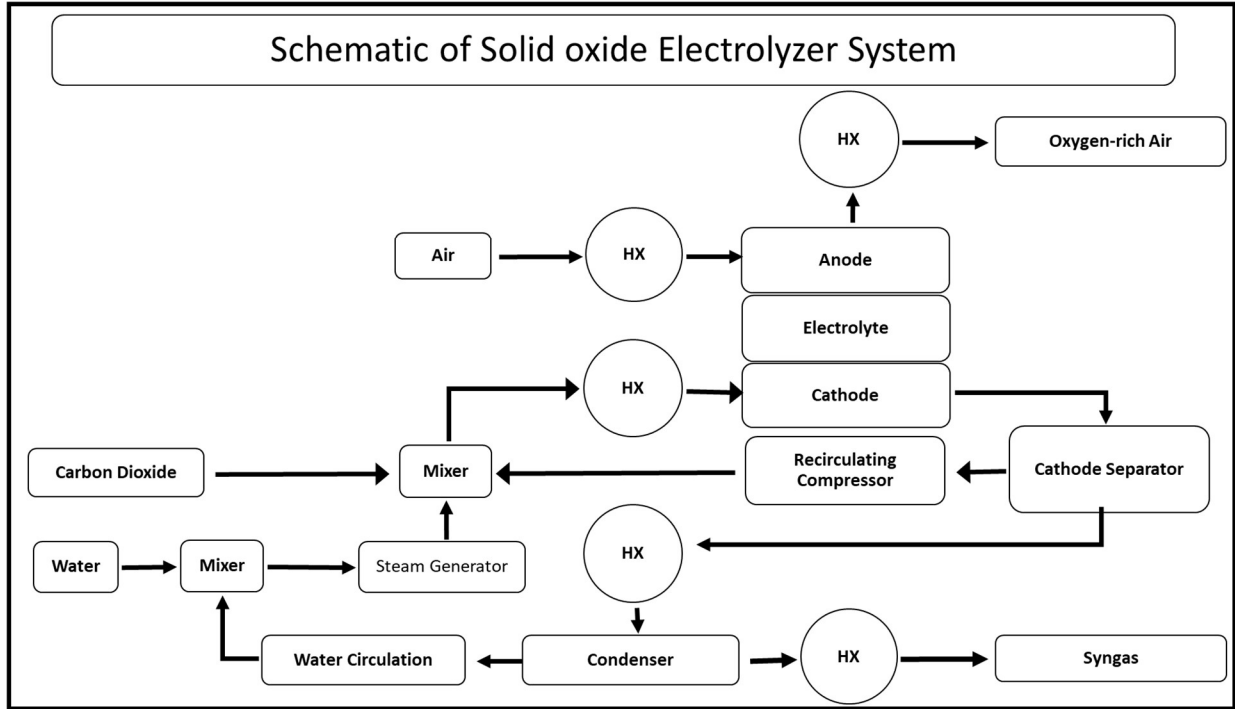
Figure 8. Schematic of SOEC.

4.1 SOEC Design

Co-electrolysis of CO_2 and H_2O have been studied for the better part of a decade as method of producing syngas using the same technology as has been used over several decades for solid oxide fuel cell (SOFC)s. Rather than producing power, SOECs consume electrical energy (and sometimes heat) to convert CO_2 and H_2O in syngas.

Solid oxide electrolysis makes use of the same material and technology as solid oxide fuel cells. A single cell in an SOEC stack is comprised of an anode, cathode, and solid electrolyte. The cathode is typically constructed of nickel and yttria stabilized zirconium and the anode is a mixture of lanthanum, strontium, and manganese oxide. When power is supplied to the SOEC, steam, and CO_2 are converted to H_2, CO, and O_2. This occurs via three reactions, namely the reverse water-gas shift and the co-electrolysis reactions:

$$CO_2 + H_2 \longleftrightarrow CO + H_2O \text{ (RWGS)} \tag{2}$$

$$CO_2 + 4e^- \rightarrow CO + \tfrac{1}{2}O_2 \text{ (Electrolysis of } CO_2) \tag{3}$$

$$H_2O + 4e^- \rightarrow H_2 + \tfrac{1}{2}O_2 \text{ (Electrolysis of } H_2O) \tag{4}$$

A model was developed in ASPEN-Plus V10 to represent product stream compositions, and heat and power requirements for an SOEC stack. Inlet mole and conversion fractions were specified to produce the required ratio of H_2/CO in the syngas for the downstream fuel production processes evaluated (~2 for methanol and ~1.2 for mixed alcohol reactors). The model was developed to simply match results from experimentally validated SOEC models from the literature (O'Brien 2009; Redissi 2013) for approximately the same syngas ratios investigated. Initially, a single-pass model, shown in Figure 9Figure 9, was generated using the same methodology as described by Redissi & Bouallou, 2013. The model introduces $CO_2/H_2O/H_2$ at a molar ratio of 45:45:10 and a temperature of 300°C to the stack. The mixture enters the stack whereupon RWGS occurs modeled as an equilibrium reactor labeled as LRWGS. A topping heater brings the mixture to the adiabatic stack operating temperature, in this case 800°C. Electrolysis occurs at thermo-neutral voltage in the stack modeled by a stoichiometric reactor (RSTOIC) using a fractional conversion from electrolysis reactions at 0.95 and 0.05 for H_2O and CO_2, respectively. RWGS again occurs at 800°C. A comparison between PNNLs model and the one created by Redissi et. al. is given in Table 5.

Figure 9. Single-pass model.
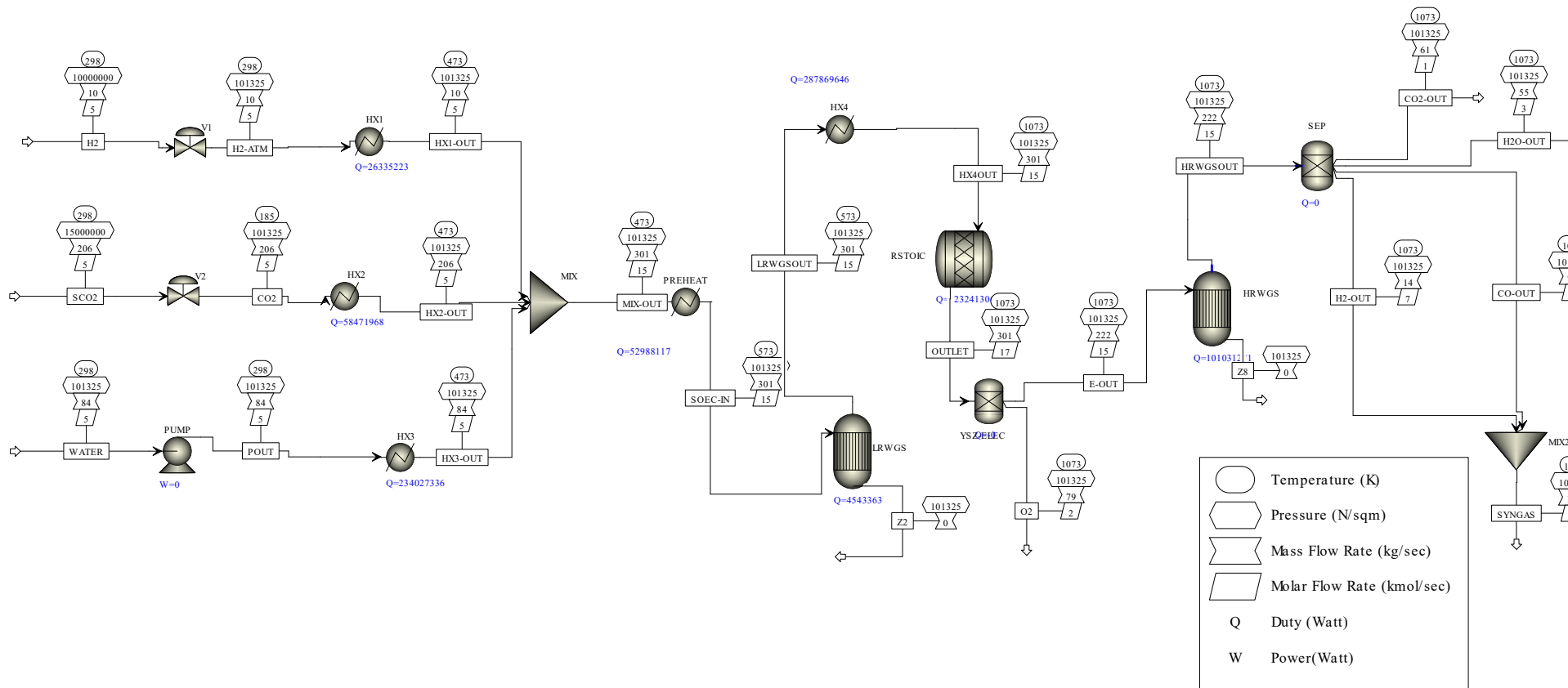
Table 5. Comparison of Redissi and PNNL SOEC models.

Outlet Gas	Redissi Model (2013)	PNNL Model for the EtOH-to-Fuels Model (Section 6)	Units
Molar Ratio	1.19	1.18	
CO	7276	7262	kg/hr
H_2	617	618	kg/hr
HX4	6061	6057	kW_t
ELECTRO (RSTOIC)	35000	32300	kW_e
HRWGS	2194	2222	kW_t

Although the Redissi model was validated experimentally, it is not representative of how a typical SOEC plant would operate. Nevertheless, the Redissi model served as a valuable first step in validating a working model for this study. Missing from the Redissi model was heat recuperation and recirculation of steam and CO_2 streams. Additionally, hydrogen is not a primary input to the stack as it is in the aforementioned model. As shown in Figure 10, the model was updated to include heat recuperation (RHX1 and RHX2), greatly reducing the power input to the topping heater. The remaining electrical heaters were replaced with steam heat exchangers available from the LWR at 260–300°C to superheat the incoming gas streams (WX1 and WX2). Approximately 30% of the hydrogen was consumed in the RWGS; this was modeled by recirculating a portion of the hydrogen product to the inlet of the stack. The final syngas ratio was 2:1.
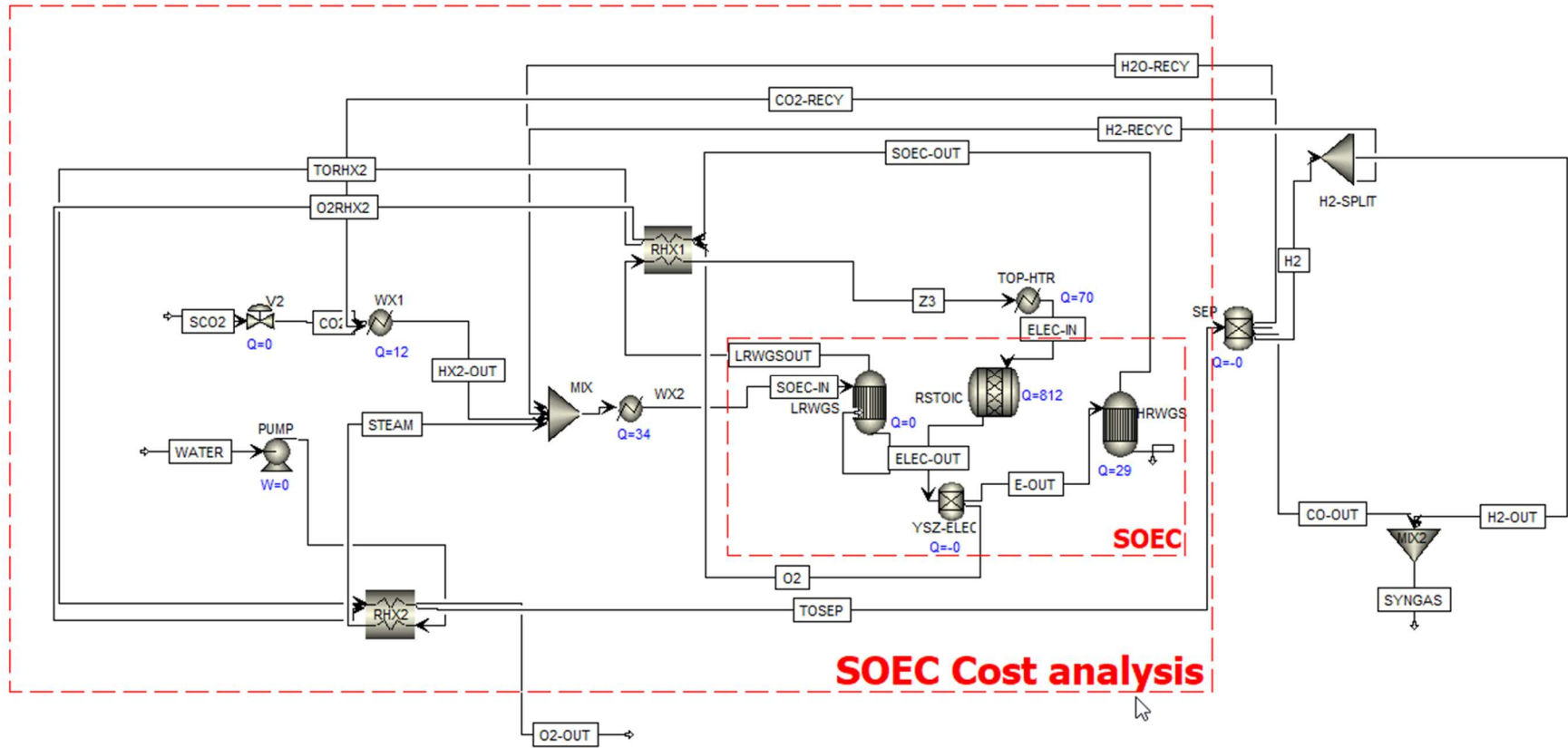
Figure 10. ASPEN-Plus model with recirculation and heat recuperation.

A comparison of a UniSim model developed by INL (O'Brien et al.2009) and the PNNL model is given in Table 6. Note that the optimal current density and area-specific resistance (ASR) from the O'Brien study were assumed for this analysis. The input and output mole fractions of the PNNL model match fairly closely with the O'Brien study. Potential reasons for the slight discrepancies are slight differences in syngas molar ratio, different modeling platforms, and the data from the O'Brien paper was read visually from a plot; therefore, the data are not stoichiometrically precise. Different SOEC operating pressures have been assumed in the literature. Redissi and Bouallou (2013) and Zhan et al. (2009) used 0.1 Pa. O'Brien et al. (Table 5), as well as Er-rbib et al (2018) and Stoots et al. (2009) used 3.5 MPa. However, a definitive relationship between pressure and process efficiency is entirely evident from the literature. A slight efficiency gain of 2.6% was reported for operating pressure of low pressure (1.6 bar) versus high pressure (5 bar) (Becker 2012).

Table 6. Comparison of O'Brien, et. al model and PNNL Model

Parameter	O'Brien, et. al. (2009)	Model for the EtOH Process (Section 6)
Inlet Composition	H_2: 24.45% CO_2: 30.23% H_2O: 45.32%	H_2: 22.2% CO_2: 26% H_2O: 51.8%
Outlet Composition	H_2: 34.57% CO:16.05% CO_2: 14.81% H_2O: 34.57%	H_2: 38% CO:13.3% CO_2: 13.2% H_2O:35.4%
Power	300 MWe	812 MWe
Operating Temperature	800°C	800°C
Operating Pressure	3.5 MPa	0.1 Mpa
Current Density[1]	0.4 A/cm^2	0.4 A/cm^2
ASR[1]	1.5 Ohm cm^2	1.5 Ohm cm^2
V_{op} (adiabatic)	1.34	1.34 (thermal neutral)
1 Optimal density and ASR selected from O'Brien		

4.2 SOEC Cost Estimation

Two different methods were used to estimate the cost of an SOEC system. One method is based on an extensive cost analysis of SOFCs, and the other is based on an area-specific cost analysis of SOECs. The SOFC method only required the total electrical power required for the stack, whereas the other method required the number of cells required as calculated from Faradays Law.

A detailed cost analysis was conducted for manufacturing, installation and operation of an SOFC for a nominal power output of 270-kW sized for ground-based distribution generation (Weimar et al. 2013). It was found that electricity costs for a mass manufactured SOFC could be $0.07/kWh based on a standard approach to manufacturing cells. A detailed study was conducted to understand the various steps required for manufacturing the units that included materials, equipment, and labor. Volumes were projected at 10,000 units/year. In addition, a sputtering approach for processing the units was considered and found to increase the performance of the stack and reduce capital costs by 33%. The cost study was based on 400-cm^2 cell area; however, the result was specific to power generation.

Given the mole flow of monatomic oxygen from the cathode to the anode as given by the model, the number of cells can be determined from Faradays Law:

$$\Delta \dot{N}_O = \frac{I_e}{2F}; \quad \text{where} \quad I_e = iA_{cell}N_{cells}$$

Where the cell area is 250 cm^2 and the current density is 0.4 A/cm^2.

Using the sputtering method, the power-specific manufacturing and installation costs were found to be \$65/kW and \$182/kW, respectively. These costs are specific to the stack and housing costs only and does not include ancillary equipment such as heat recuperators and topping heaters. These costs were used for the economic analysis of the SOEC stack for the two pathways evaluated in this study. For the MTO-based fuel pathway (278 MWe, see Section 5), total installed cost of the stack is \$68.8M and for the ethanol-based fuel pathway (425 MWe), total installed cost is \$105M.

Table 7 shows a comparison of other SOEC cost estimates from the literature compared to Weimer's method for the MTO-based model (Section 5). Giglio, et.al. (et al. 2015) estimated the area-specific costs for a stack and its enclosure to be \$1500/m^2 of SOEC active area (i.e., the cell area). Using an assumed a cell size of around 225 cm^2 (O'Brien 2009) to be acceptable given the available manufacturing techniques in 2009, and along with the current density of 0.4 amp/cm^2 and monatomic oxygen flow across the cell from the model, the total cell area required is found to be 52,986 m^2. The total cost from this approach comes to \$79.5M, for a difference of roughly 15% between the two approaches. Estimates from Buttler (2015) and Anghilante (2018) are \$93M and \$61M. The costs selected for this study (Weimar 2013) is within the range of other literature values shown.

Table 7. SOEC stack cost estimates from literature.

Source	Basis	Installed Cost (Stacks & Enclosure)	MTO Model (Section 5) SOEC Plant Cost (2019\$; Scale 278 MWe)
Weimar 2013	Fuel Cell Power	\$247/kW	\$68.8M
Giglio 2015	SOEC Area	\$1500/m^2	\$86.7M
Buttler 2015	SOEC Area	\$1755/m^2	\$101.5MM
Anghilante 2018	SOEC Power	\$219.5/kW	\$61.5M

5. SYNGAS TO FUELS VIA METHANOL-TO-OLEFINS PROCESS

5.1 Design and Modeling

The overall block flow diagram for the modeled distillate fuel production process through the MTO route is shown in Figure 11. Syngas is first generated via co-electrolysis, as described in Section 4.1. Raw syngas from the SOEC section is compressed to 420 psia. Entrained water is separated in knock-out pots prior to and between stages of compression and recycled back to the SOEC. Amine-based acid gas removal is then used to separate CO_2 from the syngas for recycle back to the SOEC. Removal of CO_2 also serves to reduce its concentration in the syngas feed to the downstream processes, thereby reducing equipment size and capital. Saturated steam at 374°F is used to regenerate the amine solvent in a mass ratio of 2:1 for steam-to-CO_2 removed (Tan et al. 2016). Single-pass conversion of CO_2 is 29% and overall carbon efficiency to CO including recycle is 98.7%.
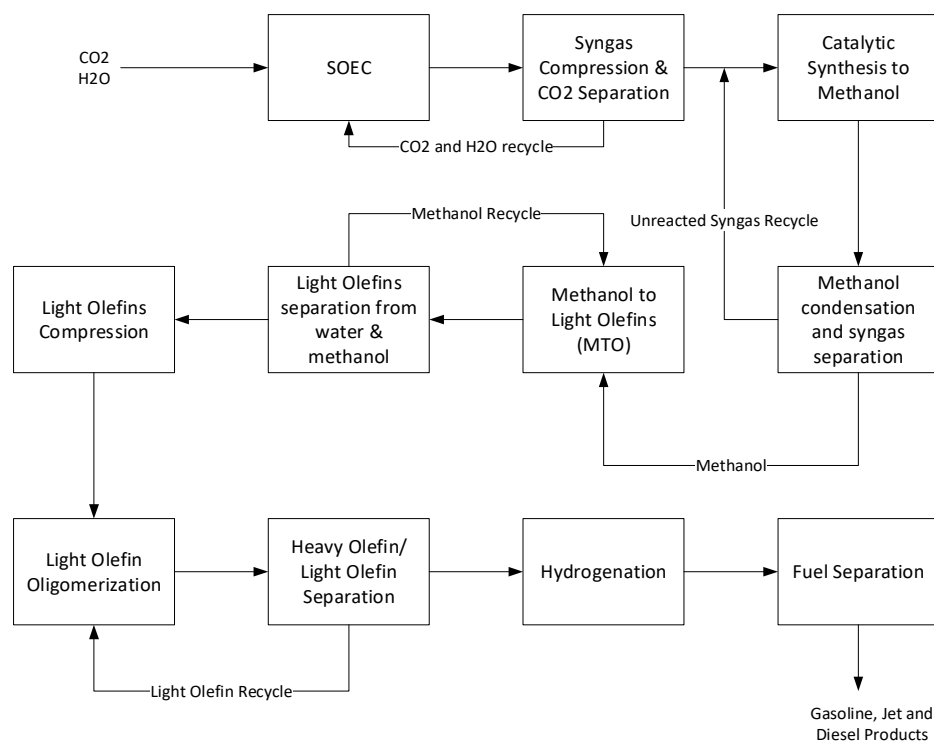
Figure 11. MTO process diagram.

Syngas is then compressed to 925 psia, mixed with unreacted syngas, heated to 440°F, and fed to the methanol reactor. Methanol synthesis occurs over a copper/zinc oxide/alumina catalyst. Process conditions for the methanol reactor are given in Table 8. Heat from the exothermic synthesis reaction is removed from the reactor via steam generation (Tan et al. 2016). Steam is used for downstream reboiler duties and for a portion of the steam needed in the CO_2 removal process. Single-pass conversion of CO is 34% and overall conversion with syngas recycle is 94%. Overall carbon efficiency of CO (and the low levels of CO_2 in syngas) to methanol is 93.2%.

Table 8. Process conditions for methanol synthesis reactor (exothermic).

Assumption	Jones, et al. (2009) Phillips, et al (2011)
Temperature, F	440
Pressure, psia	920
H_2: CO ratio	2.0
CO_2 concentration (mol%)	4%
Single-pass CO conversion	34%
Overall CO conversion	94%

Methanol is then preheated and fed to the MTO fluidized-bed reactor where ethylene, propylene, 1-butene, and 1-pentene are produced (Vora and Marker 1998). Table 9 lists the assumed process conditions and conversion efficiencies. Some lights gases and coke are formed by side reactions; the coke is burned off periodically. The effluent mixture from the MTO reactor is cross exchanged to pre-heat the inlet gas and then quenched in a direct-contact, circulating-water spray tower to separate the non-condensable gases from water generated by methanol dehydration. Residual methanol is stripped from the

quench column wastewater and recycled to the MTO reactor. The overall conversion of methanol is 99.98% with an MTO carbon efficiency of 93.4%.

Table 9. Process conditions for methanol-to-olefins (exothermic).

Assumption		Vora 1998; Gelbein 2003
Temperature, F	814	
Pressure, psia	19.7	
WHSV	1	
Single-pass Methanol conversion	99%	
Overall methanol carbon efficiency to olefins	93.4%	

The olefin-rich gas mixture goes through various separation steps and catalytic reactors to produce longer-chained hydrocarbons in the diesel-boiling range in an oligomerization step (Lilga et al. 2017). Oligomerization reactor conditions and conversions are listed in Table 10. The separation steps include a lean-oil scrubber used to recover olefins and remove methane and other paraffin by-products to prevent their accumulation in recycle streams used to increase the product yield.

Lighter olefins are recycled to increase their carbon number to the desired range of 9 to 16. Two catalysts are used to produce dimers and trimers of the reactive olefins to increase the carbon number. In the first oligomerization reactor, operated at 570°F and 302 psia, an H-Beta Zeolite catalyst is used to produce dimers and trimers of smaller (C2 to C4) olefins to C6+ olefins. The overall oligomerization reaction is exothermic; therefore, no heat is needed to drive the reaction. The assumed net conversions of ethylene, propylene and butene are 75%, 82%, and 27%, respectively. The second reactor, operated at 326°F and 140 psia, uses an Amberlyst catalyst to increase the yield of C9 to C18 compounds. The net conversion assumed for C4, C5, C6, C7, C8, and C9 olefins to their dimers are 40%, 36%, 100%, 26%, 23%, and 3%, respectively.

Table 10. Process conditions for oligomerization of olefins (exothermic).

Assumption		Reference: Dutta et al 2015; Tan et al 2016b
Temperature, F (first stage)	320	
Pressure, psia (First stage)	302	
Temperature, F (second stage)	288	
Pressure, psia (second stage)	140	
Stage 1 C2-C5 olefins C eff. to C6+ olefins	48%	
Stage 2 C2-C5 olefins C eff. to C6+ olefins	49%	
Overall C2-C5 olefins C eff. to C6+ olefins	98.6%	

Lighter olefins (in the C4 to C9 range) are recovered and recycled as part of the lean-oil scrubber feed, and the heavier olefins are hydrogenated using separately purchased hydrogen over a Pd-on-alumina hydrogenation catalyst operating at 750°F and 130 psia. The H_2 partial pressure is maintained at about 70% of total pressure to minimize coking and to completely hydrogenate all double bonds in the feed. The excess hydrogen is separated from the diesel product after condensation and knock-out pots. It is recycled to the reactor operating pressure using a pressure booster. The hydrogenated product is distilled into gasoline (9%) and diesel (91%) blend stocks in a final distillation column. Overall carbon efficiency of olefins to fuel product is 98.5%.

5.2 Performance and Economic Results

Figure 12 shows the resulting power and heat inputs and outputs for the modeled MTO process. About 92% of the power demand is for the co-electrolysis unit, with the remainder consumed for gas compressors and pumps throughout the plant. Steam is used for heating of the feed CO_2 and water to the SOEC system, for recovery of the amine solvent in the CO_2 recovery process, and for column reboilers and heating of various process streams. About 84% of the steam requirement for the integrated plant is for amine solvent recovery (shown as CO_2 Sep block below). Heat from the methanol reactor (syngas to MeOH block below) is used to produce a portion of the steam requirement, with the balance supplied by saturated steam (536°F) from the LWR. The net power and heat requirements for the fuel plant are 326 MW$_e$ and 133 MW$_t$. Electrical and thermal energy usage for syngas production for the modeled SOEC portion of the plant is 3.2 MWe/lb syngas and 19.4 MWt/lb syngas, respectively, for a H_2:CO ratio of 2.1. Note that this is only the energy usage for producing raw syngas and does not include steam used to recover amine solvent for separation of CO_2. All reactors downstream of syngas production are exothermic; therefore, no steam/heat is needed to drive these steps. For the syngas-to-methanol step, steam is generated on the shell side of the methanol reactor, capturing 31 MW of thermal energy. For the methanol-to-olefins step, 8.5 MWt of steam is generated from heat contained in the flue gas from the MTO reactor catalyst regenerator and 3.3-MWt steam is used to heat the feed stream and for the methanol recycle recovery column reboiler. For the oligomerization step, the heat generated is relatively low quality (<450°F) and is not included in the overall thermal heat generation; however, (it can be used for boiler water preheating. Distillation towers for separation of oligomers use 5.3 MWt of steam in reboilers. For the hydrotreating step, 6.3 MWt of steam is generated from the hot reactor effluent and 0.6 MWt of steam is used for reboiler heat in the diesel/naphtha fractionation column.
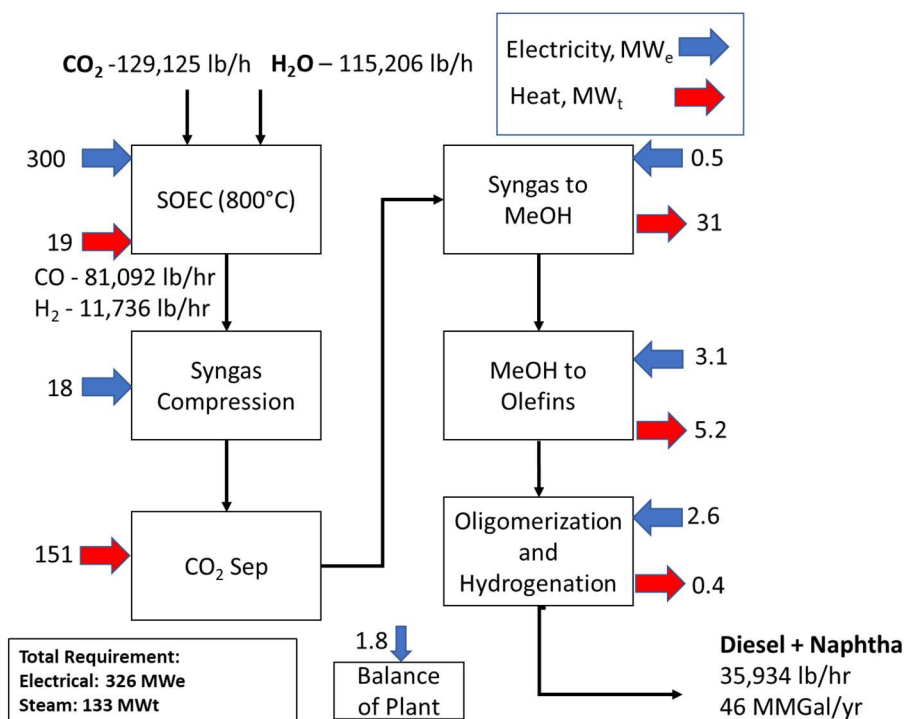
Figure 12. Electrical and thermal inputs/outputs for the synfuels via MTO process using nuclear heat.

Table 11 lists the raw material, waste disposal and utility costs assumed for estimating variable operating costs for the fuel production plant. Boardman et al. (2019) estimated the levelized cost (including capital and operating expenses) of CO_2 delivered to a hybrid LWR-methanol plant scenario including capture from an ethanol plant, compression, storage, and transportation via pipeline to the

nuclear power plant. They determined a range of $14.6 to 38.3/tonne for a 530 tonne/day capacity methanol plant. The mean of this range ($33.3/tonne) was used in this study and sensitivity analysis was conducted around this assumption. For comparison, the breakeven CO_2 sales price estimated for capture from fossil energy plants ranges from $44/tonne to $119/tonne (James et al. 2019). A range of $0 to 120/tonne CO_2 cost was investigated in the sensitivity analysis. Although clean O_2 could potentially be a co-product of the process, to be conservative, no O_2 credit is included in the analysis.

Table 11 Variable operating costs for the MTO to fuels model TEA.

Variable Operating Costs	2019 Price	Unit	Reference
Raw Materials			
CO_2	33.3	$/tonne	Boardman et al 2019
Methanol Synthesis Catalyst	12.03	$/lb	SRI PEP 2007 Yearbook
MTO Catalyst	37.14	$/lb	Gelbein 2003
1st Stage Oligomerization Catalyst	11.30	$/lb	Dutta et al. 2015
2nd Stage Oligomerization Catalyst	18.10	$/lb	Tan et al. 2016b
Hydrogenation Catalyst	59.0	$/lb	Hydrotreating Catalyst cost PNNL-13025 ($3/lb in 1987 updated to 1990 $ with PCU3251)
Amine Makeup	1.70	$/lb	Phillips et al 2007
Boiler Chemicals	3.26	$/lb	Phillips et al 2007
Cooling Tower Chemicals	1.95	$/lb	Phillips et al 2007
Hydrogen gas (for hydrotreating of oligomers)	0.95	$/lb	2020 PEP Yearbook
Waste Disposal			
Wastewater treatment	3.20	$/100 ft3	Phillips et al 2007
Utilities			
Cooling Tower makeup	236.5	¢/1000 gal	Phillips et al 2007
Process Water to SOEC	327.0	¢/1000 gal	Redissi et al 2013
Boiler Feed Water makeup	236.5	¢/1000 gal	Phillips et al 2007
Process Steam	359	¢/1000 lb steam	Boardman et al 2019, Knighton et al. 2020c
Electricity	3.0	¢/kwh	Boardman et al 2019, Knighton et al. 2020c

Table 12 gives the major performance and economic results for the synfuel plant. Also presented for comparison is the biomass case, where gasification of woody feedstock is used on the front end instead of co-electrolysis. Woody feedstock cost is assumed to be $63.23/dry ton (Hartley et al. 2019). Feedstock CO_2 and water for the SOEC are 1,549 ton/day and 331,369 gal/day, respectively. For perspective, this is about 1.75 times the daily CO_2 produced from a typical corn ethanol plant (100 million gallon/year) and about half the capacity of an Olympic-sized swimming pool, respectively. Fuel generation is about 2,900 barrel/day (BPD) of diesel and 295 BPD of naphtha (motor gasoline blendstock). The U.S. demand for diesel and gasoline in 2019 was 3.3 million BPD and 9.3 million BPD, respectively (EIA 2020b). Energy efficiency for the co-electrolysis process is similar to the gasification case, but overall carbon

efficiency is much higher due to the high selectivity of the electrolysis reactions. Feedstock cost is higher for the biomass case due to the lower carbon efficiency as compared to the co-electrolysis case. Capital costs for the syngas generation are higher for the co-electrolysis case due to the assumed cost of the SOEC stacks. However, syngas cleanup and methanol production costs are lower due to a cleaner and lower total flow of gas (fewer light ends and CO_2 than from gasification) through the compression system and reactor than results from biomass gasification. The fuel production cost breakdown is given on the bottom half of Table 12 and illustrated in Figure 13. The MFSP for the SOEC case is $4.45/gal, with electricity and steam cost making up about half of the total production cost. Optimistic and conservative cases using the lower and upper bounds respectively for estimated electricity and steam price from an LWR (2c/kWh with $2.49/1000 lb1000lb and 4c/kWh with $4.70/1000 lb1000lb steam, respectively) are included for comparison. Most of the steam requirement is for removal of CO_2 from the syngas (recovery of amine solvent). In the biomass gasification case, extra steam is generated from heat recovered from the char combustor, which supplies all heat needs for the plant as well as onsite generation of power. In the SOEC case, steam must be supplied from the LWR.

Table 12. Economic results for syngas to fuels process (all costs in 2019 $).

Flowrates	Co-electrolysis	Biomass Gasification
CO_2 Feed, lb/hr (ton/day)	129,125 (1,549)	Biomass: 183,718 (2205)
H_2O Feed, lb/hr (ton/day)	115,206 (1,381)	N/A
Diesel Blendstock, lb/hr (BPD)	32,808 (2,899)	32,835 (2,903)
Gasoline Blendstock, lb/hr (BPD)	3,126 (295)	3,129 (296)
Diesel Blendstock, mmBtu/hr (MW)	623.4 (182.7)	624.1 (183.0)
Gasoline Blendstock, mmBtu/hr (MW)	59.8 (17.5)	59.9 (17.6)
Carbon efficiency (C in synfuel/C in feed)	86.1%	32.4%
Energy efficiency (fuel)/(power+steam+H_2)	41.5%	43.0% (including input biomass)

Capital Costs, $ million		
Installed costs		
Syngas Generation	87.4	51.3
Syngas Compression and Cleanup	53.0	71.8
Methanol Production	30.6	49.0
Hydrocarbon Fuel Production	60.1	58.8
Steam Cycle / Power Gen	5.4	34.1
Balance of plant	4.8	7.7
Total installed capital cost	241.3	272.7
Indirect costs	125.2	140.9
Fixed capital investment	400.7	451.900
Total capital investment (TCI)	**422.3**	**475.2**

Operating Costs and Production Cost Breakdown				
	$ million/yr	$/gal fuel blendstock	$ million/yr	$/gal fuel blendstock
Variable operating cost				
Feedstock	15.5	0.34	45.8	0.99
Hydrogen (for hydrotreating oligomers)	4.0	0.09	3.6	0.08
Catalyst and Chemicals	3.5	0.08	4.8	0.10
Waste Disposal	0.3	0.01	6.0	0.13
Electricity	77.14	1.67	0.8	0.02
Steam	22.3	0.48		

Flowrates	Co-electrolysis		Biomass Gasification	
Fixed costs	21.9	0.47	23.7	0.51
Capital depreciation	20.0	0.43	22.6	0.49
Average income tax	5.0	0.11	5.6	0.12
Average return on investment	35.5	0.77	38.6	0.83
MFSP, $/gal fuel (91% diesel, 9% gasoline)		4.45		3.28
MFSP, $/GGE fuel (91% diesel, 9% gasoline)		4.42		3.26
MFSP, $/gal diesel blendstock		4.38		3.23
MFSP, $/gal gasoline blendstock		4.21		3.11

Economic studies from the literature for synthetic fuels production yield a wide price range. This is to be expected with the highly variable processes and technical and economic assumptions that are possible. One of the most established technologies for converting syngas to hydrocarbon fuel is via the FT route. Several pilot and demonstration tests using co-electrolysis-based syngas with FT synthesis from syngas to fuels were conducted between 2014 and 2022 in Europe (Dieterich 2020). Several groups have conducted TEAs for FT fuels via electrolysis of CO_2. Li et al. (2016) reported a range of $3.80 to 9.20/gal with a range of well-to-gate energy efficiency of 41 to 65%. Becker et al. (2012) found a range of $4.4 to 15/GGE (gasoline-gallon equivalent) for electricity price range of $0.02 to 0.14/kWh and plant capacity range of 90% to 40% and reported an overall efficiency of 51% (LHV). In a study by Fu et al. (2010), production cost ranged from $2.50 to 6.79/gal with an electricity price of $22-88MWh. Cost results from this analysis lie within the general cost range of FT fuels found in the literature.
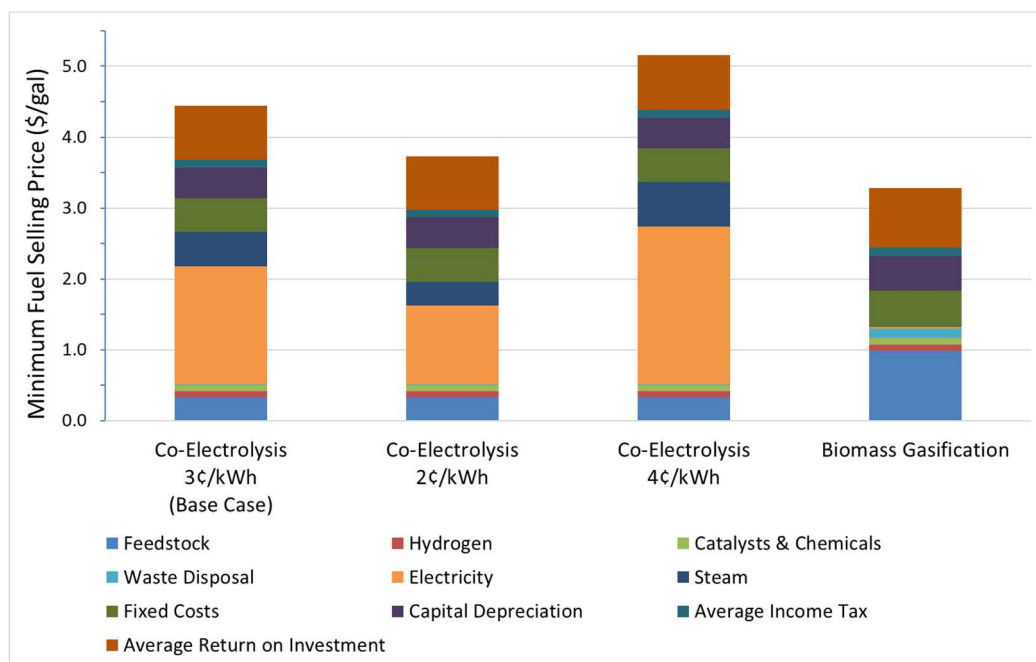

Figure 13. Fuel production cost breakdown for renewable fuel blendstock via co-electrolysis and biomass gasification and the MTO route for the conversion of the syngas to synfuels.

It is evident that the MFSP for the low-carbon fuel from this pathway fuel is higher than current market petroleum fuel prices (Figure 14). However, possible carbon credits through government

incentives or mandates are not considered in these results and could be substantial if put into place in the future. For example, credits like those granted under the Environmental Protection Agency (EPA) Renewable Fuel Standard (RFS) for biofuels may be a good first approximation for future incentives associated with CO_2-based fuels. State programs such as California's low-carbon fuel standard (LCFS) could bring additional value incentives. Renewable identification number (RIN) credits for advanced biofuel, defined as fuel with associated lifecycle greenhouse gas (GHG) emissions that is 60% lower than the petroleum baseline (e.g., gasoline and diesel), was valued in the range of $1.01 to 2.74 per gallon over the 2018–2019 time period (Figure 15, EPA 2020). Lifecycle GHG emissions for the MTO-based fuel from co-electrolysis are estimated at 13.8 g CO_2-e/MJ as shown in Table 13. This represents an 85% reduction in GHGs compared to petroleum diesel (91.8 g CO_2-e/MJ, GREET 2019). Note that this is a relatively high-level estimate that includes emissions associated with feedstock production/preparation (compression of CO_2 feed at the ethanol plant for CO_2 and woody biomass collection and preparation for the gasification case), electricity (nuclear power for the co-electrolysis case and grid power for the biomass case) and hydrogen used for hydrotreating olefins into final fuel blendstock.
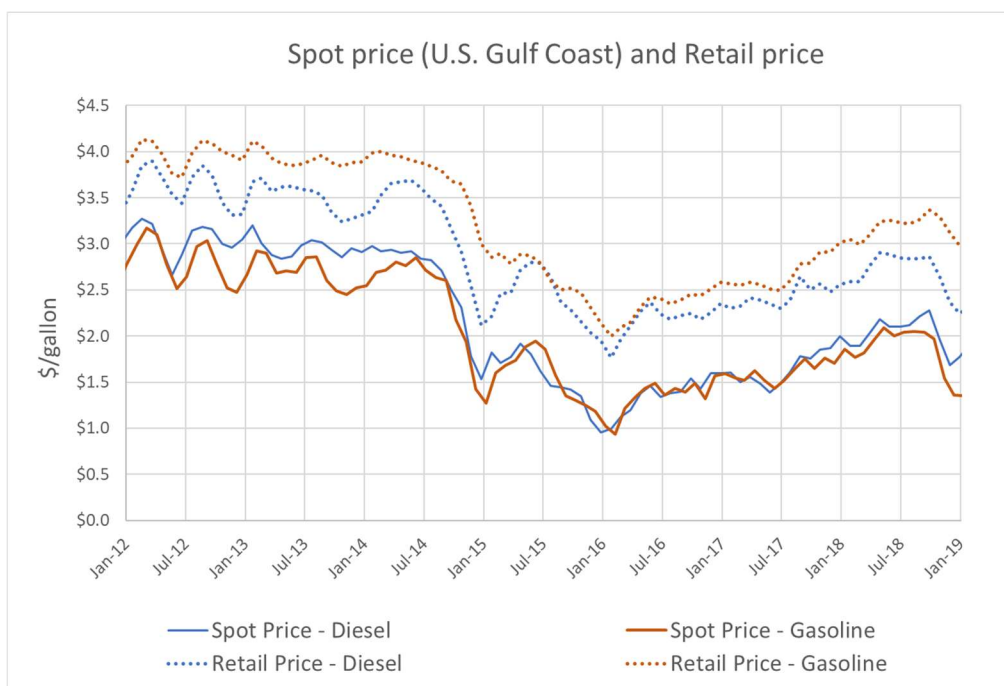

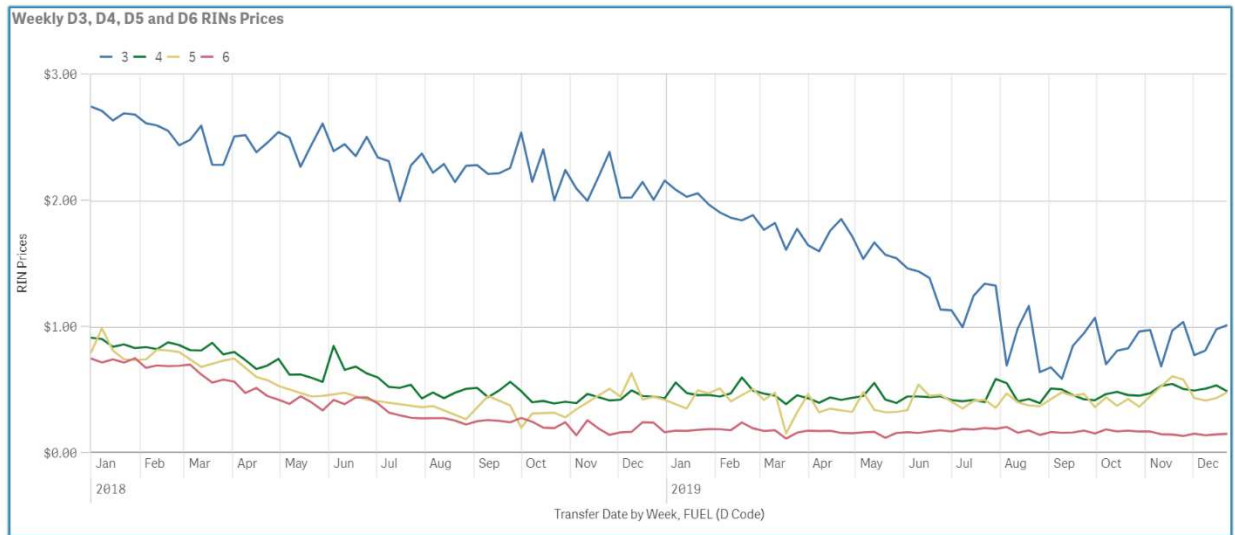Figure 14. Petroleum fuel price history (EIA 2020b).

Figure 15. Price for renewable fuel credits during 2018–2019.

Table 13. Greenhouse gas emissions calculation for fuel from the MTO-based pathway.

	Co-Electrolysis	GHG Factor and Ref	Biomass Gasification	Ref for GHG Emission Factor
Feedstock	6.2 (CO_2 compression)	157.1 kWh/tonne feed, Boardman et al. 2019; 142 kg CO_2-e/mmBtu grid mix, GREET 2019	13.9 (50/50 forest residue/clean pine)	109 kg CO_2-e/dry ton, Hartley et al. 2019
Electricity	3.6 (nuclear)	2.4 kg CO_2-e/mmBtu, GREET 2019	0.7 (grid mix)	142 kg CO_2-e/mmBtu, GREET 2019
Hydrogen for Hydrotreating	3.9	105 kg CO_2-e/mmBtu, GREET 2019	3.9	105 kg CO_2-e/mmBtu, GREET 2019
Total*	13.8		18.5	
Reduction from Petroleum Diesel	85%		80%	

*Does not include contribution of chemicals and catalysts consumption.

Sensitivity analysis (Figure 16) investigating variable CO_2 cost, CO_2 credits, and electricity price shows that with optimal CO_2 and electricity prices and inclusion of carbon credits through incentives or mandates could make this process more cost competitive with petroleum fuels. With a hypothetical carbon tax of \$100/tonne CO_2 the MFSP is reduced to ~\$3.75/gallon. An RFS credit would further aid in competitiveness of fuels produced via this route. Carbon sequestered into fuel minus CO_2 emitted in process off gas combustion and embodied GHGs for H_2 requirement for hydrotreating of the final fuel was estimated. The carbon credit reduces MFSP by approximately \$0.86/gal, somewhat less than the average 2018/2019 D3 advanced biofuel RIN price of \$1.88. This analysis indicates that fuel from this pathway could be more competitive with petroleum prices with a combination of lower electricity price, lower CO_2 price, and qualifying carbon credits. If a D3 RIN-type price could be applied, competitiveness could be further improved.

Also, there are innovative cryogenic carbon capture (CCC) processes that could have significant impact on the viability of an LWR/synfuels plant using methanol as intermediate. Further, the refrigerant used in the CCC process could be produced using LWR energy. The synergies of the LWR with the CCC process and techno-economic modeling of the CCC process will be explored in detail in future studies.
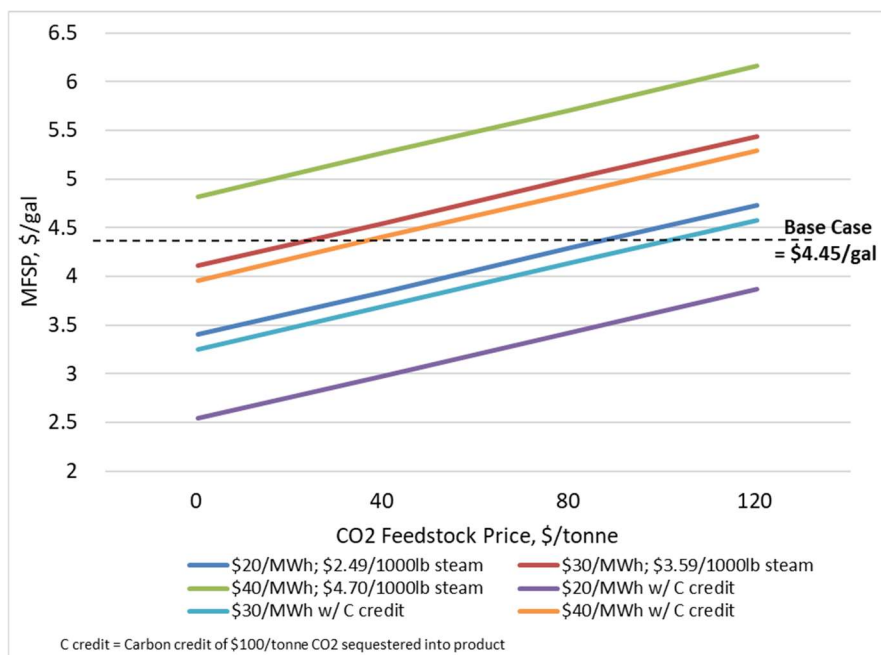
Figure 16. Sensitivity of MFSP for MTO fuel to CO_2 and electricity price and considering the potential impact of a carbon credit.

The scale of the processing plant is also an important cost driver for any chemical process due to manufacturing economies of scale. Economies of scale are realized for most equipment, stemming from the fact that the surface area to volume ratio for cylinders and spheres decreases as capacity (volume) is increased. Sensitivity analysis varying plant scale for the co-electrolysis with MTO fuel process was conducted, the results of which are shown in Figure 17. At a scale of half the base case (326 MWe; 133 MWt), production cost increases by 9%. At a scale 10 times larger than the base case, production cost is reduced to about $3.8/gal. Scaleup of the plant up to the entire electrical output of a general 1-GWe LWR of fuel production would result in about 40 cents/gal cost savings. Note that a scaling factor of 1 is assumed for the SOEC stack; therefore, no benefit is gained for this portion of the capital cost. This is because SOEC stacks are built up in modules. To increase the scale of an electrolysis system, more modules of the same size are added to the system.
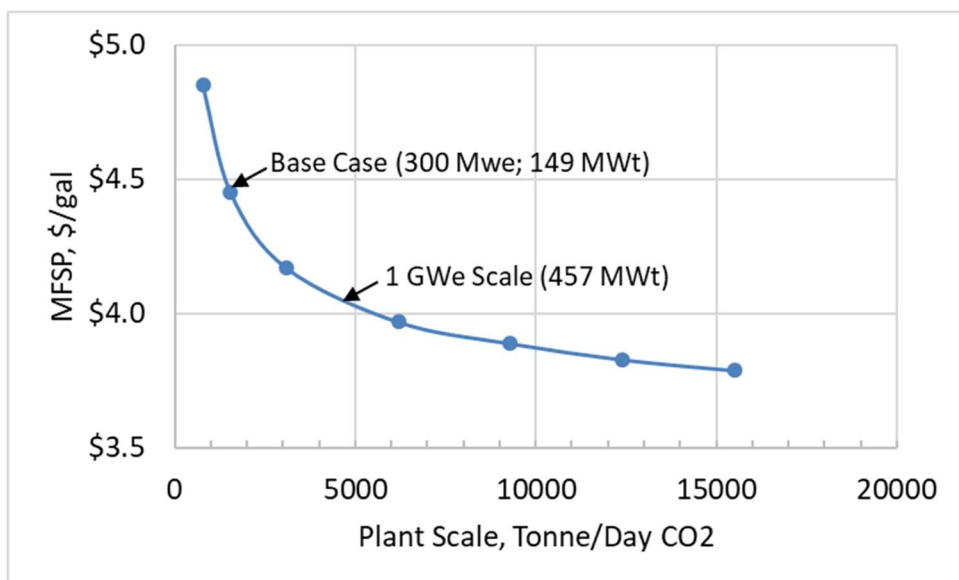
Figure 17. Sensitivity of MFSP to plant scale for the co-electrolysis and MTO fuel process.

6. SYNGAS TO FUELS VIA MIXED ALCOHOLS PROCESS

6.1 Design and Modeling

The overall block flow diagram for the modeled fuel production process through the ethanol route is shown in Figure 18. Similar to MTO pathway, unreacted CO_2 is captured using an amine-based solvent while water is separated in a knock-out tank prior to recycling back to the SOEC. The syngas from the amine absorption column is compressed to 3,000 psi (207 bar) using a five-stage centrifugal compressor with inter-stage cooling. Single-pass conversion of CO_2 is 34% and overall carbon efficiency to CO including recycle is 98.5%.
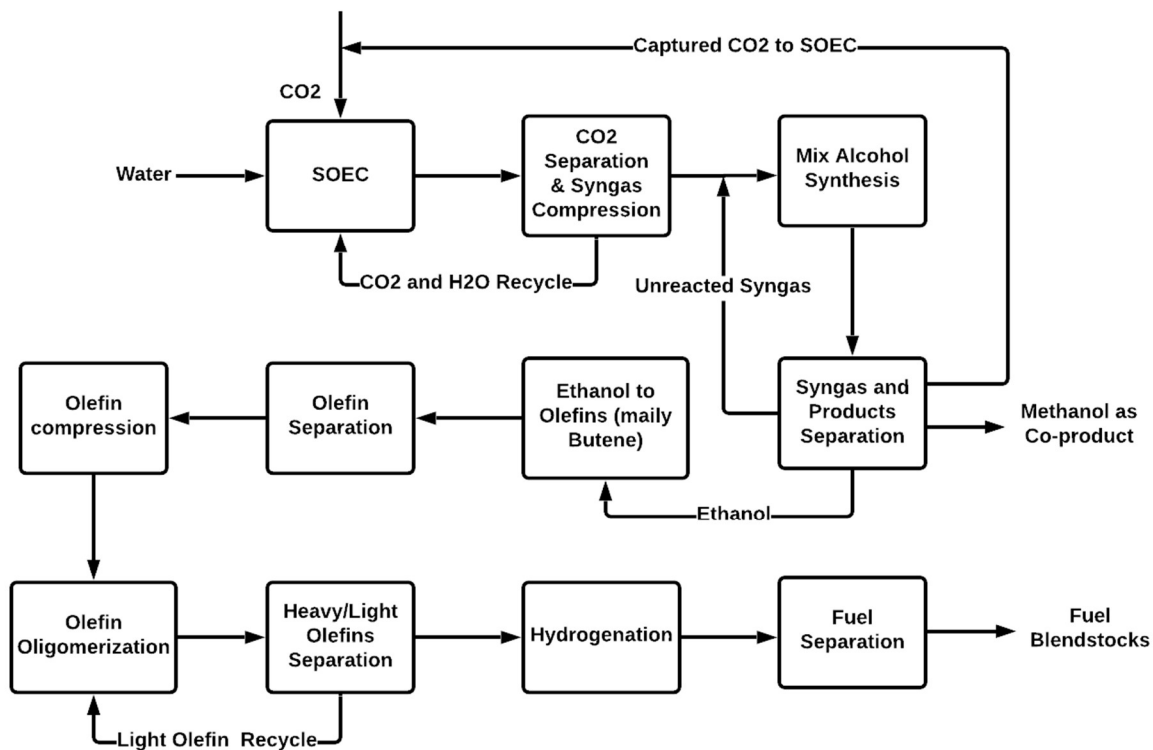
Figure 18. Syngas to fuel via EtOH pathway.

The compressed syngas is then mixed with recycled syngas and methanol and preheated to 595°F (313°C) before entering the alcohol synthesis reactor. Within the alcohol reactor, the syngas contacts a metal-sulfide catalyst to product methanol and ethanol with a small number of light hydrocarbons (e.g., methane, ethane, and propane). The optimal H_2/CO ratio for the mixed alcohol reaction with metal-sulfide catalysts is 1.5, which is less than the stoichiometry ratio in the desired product due to the employed catalyst have strong water-gas-shift reaction ($CO + H_2O \rightarrow CO_2 + H_2$) (Dutta et al. 2011). For this reason, an H_2:CO ratio of less than 2.0 is preferred for this design. Heat must be removed from the reactors because the synthesis reaction is exothermic. Reactor temperature is controlled at 611°F by generating saturated steam at 681 psia. Table 14

Table 14 lists process conditions and performance for the mixed alcohol reactor. The reactor effluent is cooled in a series of exchangers to condense the alcohols, which are then separated from unreacted syngas in a flash vessel. The gas stream goes through a set of acid gas removal units to separate the unreacted syngas from undesired gases before recycling back to the mixed alcohol reactor. Also acid gas removal unit here is to capture CO_2 to be used as carbon source for SOEC (Samavati et al. 2018; Becker et al. 2012). About 10% gas is purged to prevent excessive buildup of inert gases (e.g., light hydrocarbons) for heat generation. The methanol/ethanol mixture is separated in a distillation column, where the ethanol and some propanol is recovered from the bottoms containing 99% of the ethanol fed to the column, and the top stream is further processed to produce 99.85 wt% industrial-grade methanol. Single-pass conversion of CO is 29% and overall conversion with syngas recycle is 79%. Overall carbon efficiencies of CO to ethanol and methanol are 45%, and 5%, respectively.

Table 14. Process conditions for mixed alcohol synthesis

Assumptions	Values	Reference
Temperature, F	611	Dutta et al. 2011
Pressure, psia	3000	
H_2: CO ratio	1.5	
CO_2 concentration (mol%)	15%	
Sulfur (ppm)	70	
Single-pass CO conversion	29%	
Overall CO conversion	79%	
CO selectivity to ethanol	63%	

Ethanol is then preheated and co-fed with hydrogen (5 mol% H_2 in the feeding) to multiple reactors in parallel for producing 1-butene with a small amount of ethylene, propylene, acetaldehyde, diethyl ether and alkane. Some coke is formed by side reactions; the coke is burned off by taking each reactor into regeneration mode on a continuous cycle. The total olefin selectivity over Ag/ZrO2/SiO2 catalyst is 88% with 1-butene selectivity of 65% based on the PNNL experimental data. The effluent mixture from the olefin reactor (mainly butene) is cross exchanged to pre-heat the inlet gas and then quenched in a direct-contact, circulating-water spray tower to separate the non-condensable gases from water generated by ethanol dehydration. Conversion of ethanol is 100% and carbon efficiency of ethanol to olefins is 95.5%.

The butene-rich gas mixture is compressed to 340 psi (23 bar) and heated to 437°F (225°C) before entering the first oligomerization reactor to produce C4, C6 and C8 oligomerization products. Ni on a silicoaluminate catalyst can be used in the first oligomerization stage. The product from the first oligomerization stage is introduced to the second oligomerization reactor to increase the yield of C9 to C16 compounds. The net conversion assumed for C2, C3, C4, C6, and C8 olefins to their dimers are 88%, 77%, 88%, 20%, and 50%, respectively (Lilga et al. 2016).

Lighter olefins (in the C4 to C9 range) are separated from the heavier olefins via flash evaporation and distillation. The light olefin is recycled back to the first oligomerization reactor while the heavier olefins are hydrogenated using purchased hydrogen over a Pd-on-alumina hydrogenation catalyst operating at 700°F (371°C) and 300 psi (20 bar). The H_2 partial pressure in the hydrotreater is maintained at about 70% of total pressure to minimize coking and to completely hydrogenate all double bonds in the feed. The excess hydrogen is separated from the diesel product after condensation and knock-out pots. It is recycled to the reactor operating pressure using a pressure booster. The hydrogenated product is distilled into diesel blendstocks in a final distillation column. Overall carbon efficiency of olefins to fuel product is 97.3%.

6.2 Performance and Economic Results

Figure 19 shows the resulting power and heat inputs and outputs for the modeled EtOH pathway. The net power requirement for the plant is 468 MWe and the steam requirement is 66 MWt. The co-electrolysis unit consume about 91% of the electricity demand and the rest 9% is mainly for gas compressors and pumps throughout the plant. Steam is needed for heating of the feed CO_2 and water to the SOEC system, amine solvent regeneration in the CO_2 separation (shown as CO_2 Sep block below) and heating demand for downstream. As shown in Figure 19

, the most significant steam requirement for the integrated plant is for amine recovery in the CO_2 separation. When comparing to MTO pathway, the steam requirement from LWR for the integrated plant is only 44% of steam demand for the MTO pathway. This is because the heat recovered from the mixed alcohol reactor and from the offgas gas combustor is substantial. Electrical and thermal energy usage for syngas production for the modeled SOEC system is 2.5 kWe/lb syngas and 0.2 kWt/lb syngas, respectively for a H_2:CO ratio of 1.2. Similar to the methanol pathway (Section 6.1), all reactors downstream of syngas production are exothermic; therefore, no steam/heat is needed to drive these steps.
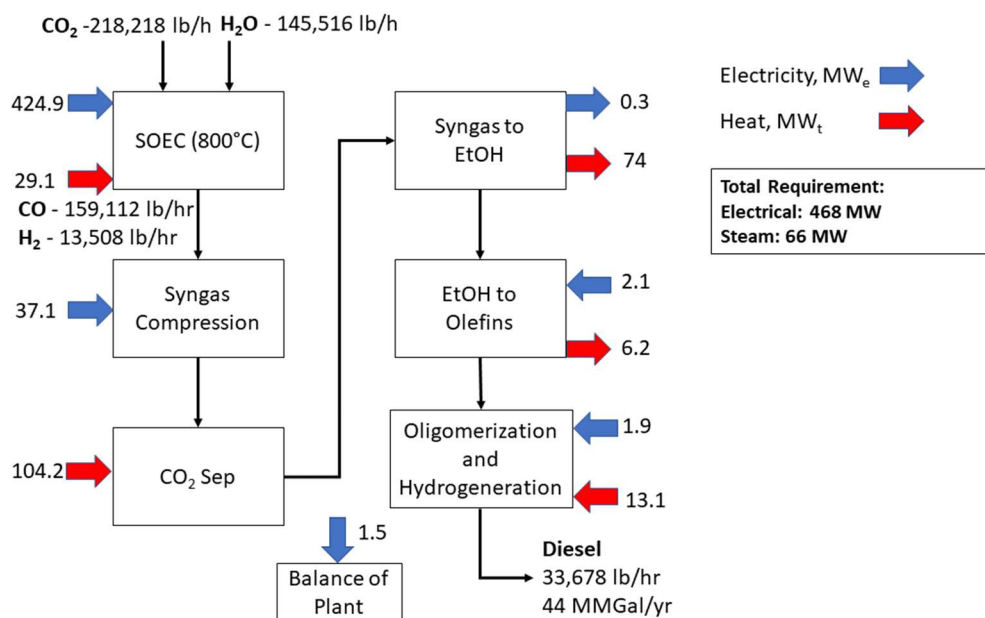
Figure 19. Heat and power inputs/outputs for the syngas-to-fuel process using SOEC.

Figure 20 shows the carbon distribution analysis result for this pathway, which reveals that the carbon loss in purge gas of mix alcohol synthesis is ~34.9% of the feeding carbon. As mentioned before, the purge stream is to control the inert gases concentration in the mixed alcohol synthesis (MAS) reactor feeding so that lower the MAS reactor cost. The carbon loss of MAS includes unconverted syngas (6.5%), produced alkane (14.6%), rejected CO_2 (13.8%). Such considerable carbon loss can be partly attributed to the low CO conversion and low selectivity in the MAS reactor. In the Co-electrolysis case, this purge stream is used as fuel to produce heat for this process. But for biomass case, this stream can be recycled back to tar reformer to recover the carbon in the syngas and light alkane. This is the main reason that biomass case shows better carbon efficiency, as shown in Table 14. To improve the carbon efficiency, 70% CO_2 is captured to be recycled back to SOEC subsystem. The overall carbon efficiency for fuel

32

product is 41.4%, nearly 50% of the carbon efficiency of MTO pathway. A higher selectivity catalyst can further increase the carbon efficiency and decrease the cost for this pathway.
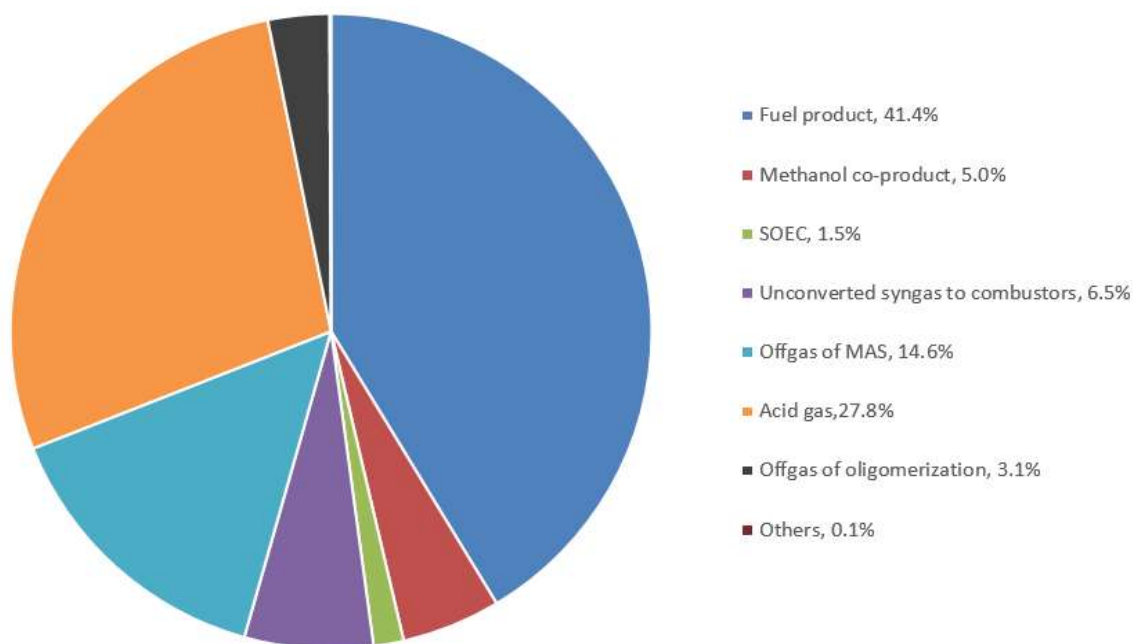
Figure 20. Carbon distribution for SOEC syngas to fuel via EtOH pathway.

Table 15 lists the catalyst cost for estimating variable operating costs for this pathway and please refer to Table 8 for CO_2, waste disposal, and utility costs.

Table 15. Variable operating costs for the MTO to fuels model TEA.

Catalyst and Chemicals	2019 Price	Unit	Reference
Mixed Alcohol Catalyst	42.67	$/lb	Dutta et al. 2011
Butene (C4) Catalyst	1.15	$/lb	2014 PEP Yearbook for Acetic Acid using a silver gauze catalyst
Oligomerization Catalyst	18.10	$/lb	Tan et al. 2016b
Hydrogenation Catalyst	59.03	$/lb	Hydrotreating Catalyst cost PNNL-13025 ($3/lb in 1987 updated to 1990 $ with PCU3251)
DEPG makeup	95.59	$/million lb acid gas removed	Dutta et al. 2011
Selective amine makeup	21.17	$/million lb acid gas removed	Dutta et al. 2011
Hydrogen (for hydrotreating oligomers)	0.95	$/lb	2020 PEP Yearbook
Methanol	0.38	¢/gal	2020 PEP Yearbook

Table 16 gives the major economic results for fuel plant via EtOH pathway. Also presented for comparison is the biomass case, where gasification of woody feedstock is used on the front end instead of co-electrolysis. The fuel production cost breakdown for each case is given in Figure 21. The fuel MFSP for the pathway is \$6.13/gal. The electricity and utility (mainly steam) and CO_2 feeding for the SOEC are the major cost drivers. Electricity and steam cost contribute 45% of the MFSP while CO_2 feeding contributes 10% of the MFSP. Compared to MFSP of the MTO pathway, the MFSP for EtOH pathway increased by \$1.68/gal due to the lower carbon efficiency as explained earlier thus higher capital cost associated with large SOEC and expensive equipment cost for acid gas removal system.

Table 16. Economic results for syngas to fuels process (all costs in 2019 $).

Flowrates	Co-electrolysis	Biomass Gasification
CO_2 Feed, lb/hr (ton/day)	218,218 (2,619)	Biomass: 183,718 (2,205)
H_2O Feed, lb/hr (ton/day)	145,165 (1,742)	N/A
Fuel, lb/hr (BPD)	33678 (2,873)	33,672 (2,869)
Fuel, mmBtu/hr (MW)	637.2 (186.7)	636.2 (186.4)
Carbon efficiency (C in synfuel/C in feed)	41.4%	30.4%
Energy efficiency (fuel)/(power+steam+H_2+natural gas)	32.7%	41.5% (including input biomass)

Capital Costs, $ million		
Installed costs		
Syngas Generation	115.6	49.9
Syngas Compression and Cleanup	57.5	110.8
Methanol Production	80.0	84.5
Hydrocarbon Fuel Production	27.0	25.6
Steam Cycle / Power Gen	3.3	33.5
Balance of plant	7.4	9.2
Total installed capital cost	290.8	312.5
Indirect costs	147.7	159.0
Fixed capital investment	472.7	508.7
Total capital investment (TCI)	497.9	535.7

Operating Costs				
	$ million/yr	$/gal	$ million/yr	$/gal
Variable operating cost				
Feedstock	27.9	0.64	45.8	1.10
Hydrogen	1.9	0.04	1.5	0.04
Catalyst and Chemicals	19.8	0.45	5.2	0.13
Waste Disposal	0.0	0.00	2.5	0.06
Electricity, Steam and Other Utilities	119.1	2.73	10.4	0.25
Co-product credits	15.8	0.36		
Fixed costs	-13.9	-0.32	0.00	0.00
Capital depreciation	24.6	0.56	25.9	0.62
Average income tax	23.6	0.54	25.4	0.61
Average return on investment	5.9	0.14	6.3	0.15
MFSP, $/gal biocrude		42.7	0.98	44.1
MFSP, $/GGE biocrude		5.86		3.82
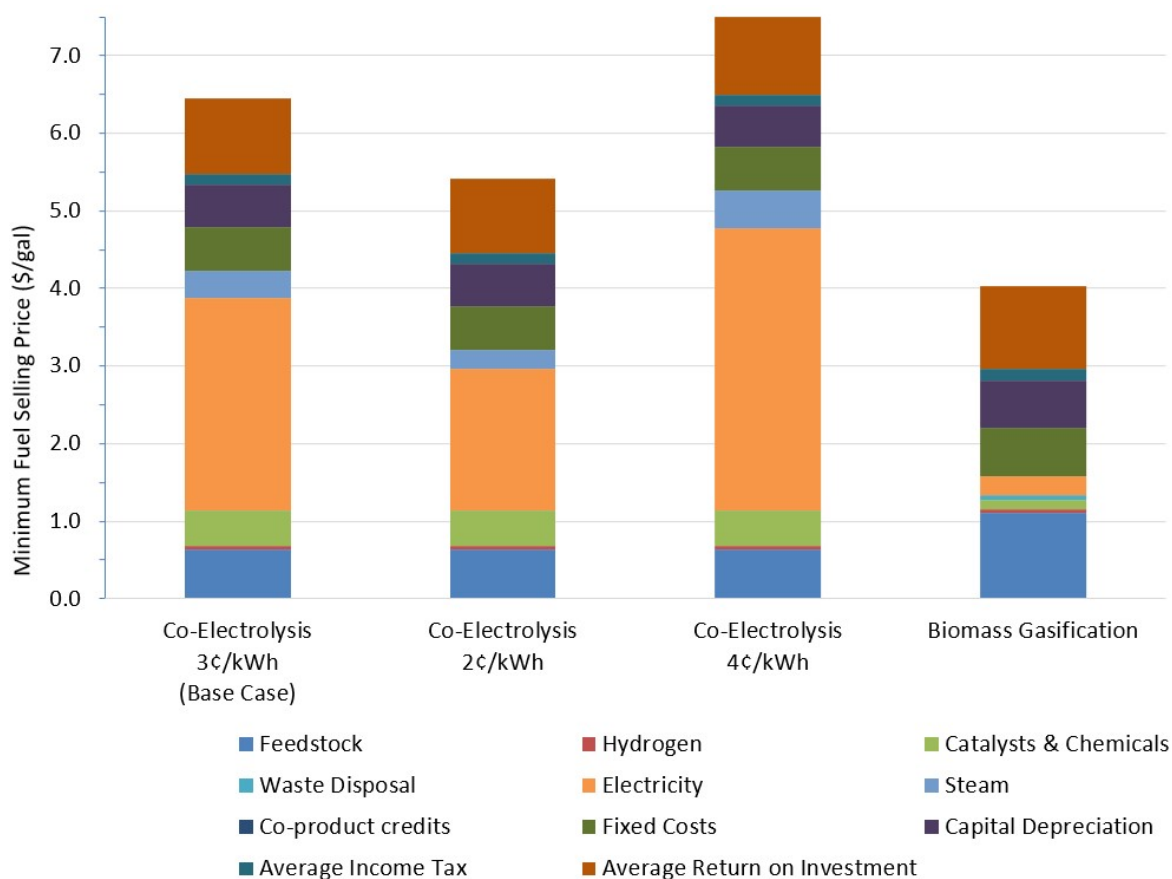
Figure 21. Fuel production cost breakdown for renewable fuel blendstock via the EtOH route.

A similar sensitivity study investigating variable CO_2 cost and electricity price is presented in Figure 22. Cases with and without a $100/ton carbon credit are included to investigate the possible impact of government mandates or incentives. The carbon credit reduces MFSP by $1.35/gal for cases with the same electricity and steam price, which is higher than MFSP reduction of MTO pathway. This is because more CO_2 is captured/installed for per mass fuel product via EtOH pathway. Table 17Table 17 shows GHG emissions for co-electrolysis and biomass via EtOH pathway. It is found that lifecycle GHG emissions (18.7 g CO_2-e/MJ for fuel from EtOH pathway) are reduced by 80% compared to the petroleum diesel (91.8 g CO_2-e/MJ, GREET 2019), which shows higher GHG emission reduction than biomass case. Note that the calculated GHG emissions including emissions for CO_2 compression, nuclear power and hydrogen inputs (GREET 2019).
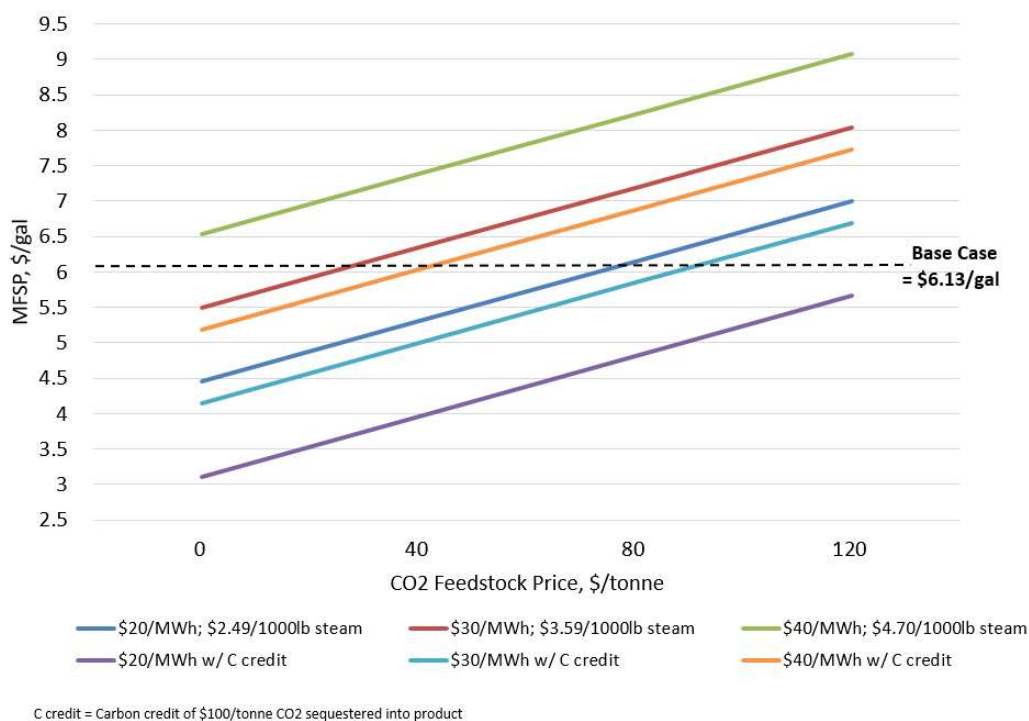
Figure 22. Sensitivity of MFSP for EtOH fuel to CO_2 and electricity price and considering the potential impact of a carbon credit.

Table 17. Greenhouse gas emissions calculation for synfuel via EtOH pathway.

	Co-Electrolysis	GHG Factor and Ref	Biomass Gasification	Ref for GHG Emission Factor
Feedstock	11.2 (CO_2 compression)	157.1 kWh/tonne feed, Boardman et al. 2019; 142 kg CO_2-e/mmBtu grid mix, GREET 2019	15.0 (50/50 forest residue/clean pine)	109 kg CO_2-e/dry ton, Hartley et al. 2019
Electricity	5.6 (nuclear)	2.4 kg CO_2-e/mmBtu, GREET 2019	12.6 (grid mix)	142 kg CO_2-e/mmBtu, GREET 2019
Hydrogen for Hydrotreating	1.9	105 kg CO_2-e/mmBtu, GREET 2019	1.9	105 kg CO_2-e/mmBtu, GREET 2019
Total*	18.7		29.5	
Reduction from Petroleum Diesel	80%		68%	

*Does not include contribution of chemicals and catalysts consumption.

7. LOW-CARBON FUEL CREDITS APPLICABLE TO SYNTHETIC FUELS PRODUCTION

A recent report (Knighton 2020b) studied the framework of options that can incentivize nuclear power plant operations by providing credits for the low-carbon grid power and non-electric products that may be produced from nuclear power, similar to renewable energy credits (REC). Some conclusions from this report are highlighted below as they apply to synthetic fuel produced by coupling with low-carbon nuclear energy. The report highlighted the status quo of the EPA RFS, carbon tax/credit systems, LCFS of California, the new green hydrogen standard in New York as well as other possible future frameworks that may incentivize nuclear energy operators and downstream industry employing low-carbon electricity and non-electric products. More generally these credits can be termed zero emissions credits (ZEC), including renewables and nuclear energy.

For example, electricity, hydrogen, and products produced from hydrogen such as steel and ammonia could create ZECs or "low-carbon" green energy credits that can be used by obligated industry entities needing to reduce their carbon footprint. Green steel produced from hydrogen using nuclear energy could qualify for very large (~$150/tonne) carbon credits in the European export markets. It is conceivable that synthetic fuel produced using low-carbon nuclear energy and feedstock CO_2 that would otherwise be exhausted to the atmosphere from natural gas combined cycle or ethanol plants, could be included in these existing and future national and state "ZEC" programs.

Other reports completed by the DOE LWRS program have highlighted the vast and diverse markets for non-electric products that can be produced using nuclear energy (Knighton 2020a, Hu 2019, Frick 2019). The current report has supplemented these studies by providing a first look into two possible pathways for producing synthetic transportation fuels by coupling with low-carbon nuclear energy.

Nuclear energy is a large portion of the low-carbon generation mix in the U.S. There is precedent for nuclear energy being included in existing and proposed clean energy frameworks and legislation (New York and Illinois Clean Energy Standards for electricity generated from nuclear energy, California LCFS for transportation fuels, and New York curtailed hydrogen credits). Electricity, hydrogen, and synthetic fuels produced from nuclear energy can be considered low carbon and comparable to renewable energy such as solar and wind even after the entire life cycle is considered (including uranium mining, fuel manufacture, plant construction, etc.).

Retiring nuclear plants and not valuing this low-carbon energy with the commensurate credits given to renewable energy may lead to drastic increases in carbon emissions (from substitute baseload plants such as NGCC) at a time when decreases in carbon emissions are being sought, which would be contrary to the goals of decarbonization.

Important points to consider related to possible future low-carbon / zero emissions credit legislation:

- The retention of nuclear power generation is critical to achieving federal and states' decarbonization goals across multiple energy sectors

- Producing hydrogen and other products such as synthetic fuels from nuclear power, especially at low demand periods, increases the capacity utilization factor of NPPs, which can improve the economics of their operation

- The contribution of nuclear power to zero-carbon power markets can be extended further to serve other energy sectors such as transportation, as well as building and industrial heat demand, thus contributing to the goals of decarbonization across multiple energy sectors.

8. CONCLUSIONS

Detailed process modeling and techno-economic analyses have been conducted for two potential power-to-fuels processing routes for integration with steam and electricity from an LWR, namely using

co-electrolysis to produce syngas and then either using a methanol to synfuels pathway or an ethanol to synfuels pathway. The methanol to synfuels pathway appears to be more economical than the ethanol pathway. The models reflect a conceptual fuel production plant nearby or co-located with an LWR that uses CO_2 offgas from an ethanol production plant in the region as feedstock to a co-electrolysis SOEC system for syngas production followed by conversion of syngas to oxygenate intermediate (methanol or ethanol), and final diesel and naphtha blendstock finishing. It is important to note that these results reflect a low-level cost estimate (e.g., "study estimate") due to uncertainties around the costs and performance of the SOEC and other downstream steps that are still in development. It should also be emphasized that these costs reflect n^{th}-plant economics (see Section 3.1), which takes advantage of assumed technology improvements at a future time after which several plants have been built and are operating. As such, they do not include additional expenses that can be expected with building a first-of-a-kind plant, such as longer startup times and large project contingencies.

Using the base case assumptions of \$33.3/tonne CO_2 cost and \$30/MWh electricity cost, the modeled MFSP for the methanol-based route to fuel is estimated to be \$4.45/gal for a fuel production plant of 3,194 BPD capacity (91% diesel and 9% motor gasoline blendstock). Electricity and steam consumption from the LWR for fuel production are 326 MWe and 133 MWt, respectively. Production costs for the co-electrolysis routes using these above assumptions leads to a cost about 40% higher than using wood gasification for syngas production, primarily due to electricity usage and steam usage for co-electrolysis and CO_2 separation (heat from gasification supplies all the steam and almost all of the power needs for the gasification plant). The assumptions used in the base case are conservative and do not account for improving technologies in the areas of CCC.

This analysis highlights the opportunities available to improve the economics of synfuel production through co-electrolysis routes to be competitive with biomass gasification and petroleum refining:

1. Potential reduction in cost of CO_2 feedstock from the \$33.3/tonne CO_2 using innovative CCC processes to capture carbon from ethanol plants and investigation into the cost and synergies of other sources such as NGCC plants.

2. Potential CO_2 credits and/or LCFS program credits (California as a first example) or RFS credits under the EPA.

3. There is considerable opportunity in producing various high-value synthetic chemicals, which could be the subject of future techno-economic studies. This report investigated only two possible synthetic fuels routes.

Overall carbon efficiency of CO_2 to fuel is about 2.75 times higher than the biomass gasification case due to the high efficiency of the co-electrolysis process. Thermal efficiencies of the co-electrolysis and biomass cases are similar. Sensitivity analysis shows that a combination of lower electricity and steam price (\$20/MWh and \$2.49/1000 lb1000lb steam), low CO_2 price (\$15-50/ton), and inclusion of carbon credit of a \$100/tonne can make the process more competitive with petroleum fuels. GHG analysis indicates that the fuels have 85% less GHG emissions than petroleum fuels (diesel) and could conceivably qualify for the highest RIN credit if approved by the EPA. This, along with additional state incentives could bring an even higher potential credit for the fuel.

Economic studies from the literature for synthetic fuels production yield a wide price range. This is to be expected with the highly variable processes and technical and economic assumptions that are possible. One of the most established technologies for converting syngas to hydrocarbon fuel is via the FT route. Several pilot and demonstration tests using co-electrolysis-based syngas with FT synthesis from syngas to fuels were conducted between 2014 and 2022 in Europe (Dieterich 2020). Several groups have conducted TEAs for FT fuels via electrolysis of CO_2. Li et al. (2016) reported a range of \$3.80 to 9.20/gal with a range of well-to-gate energy efficiency of 41 to 65%. Becker et al. (2012) found a range of \$4.4 to 15/GGE (gasoline-gallon equivalent) for electricity price range of \$0.02 to 0.14/kWh and plant capacity

range of 90% to 40% and reported an overall efficiency of 51% (LHV). In a study by Fu et al. (2010), production cost ranged from $2.50 to 6.79/gal with an electricity price of $22 to 88MWh. Cost results from this analysis lie within the general cost range of FT fuels found in the literature. The variability of these study results presents a future opportunity for fine tuning modeling and results to determine the most viable routes for producing synthetic fuels when coupled with an LWR.

The base case modeled MFSP for the ethanol-based fuel route is $6.13/gal. This option has a higher production cost than the methanol route primarily because the syngas to ethanol process is only half as carbon efficient as the syngas-to-methanol process. This leads to about twice the demand for CO_2 feedstock and electricity in the SOEC. Compared to the biomass gasification case, production cost is about 52% higher. Overall carbon efficiency to fuel is 41.4% (46.4% including methanol co-product), about 40% higher than the biomass gasification case. Thermal efficiency is 34.7%, about 21% lower than the gasification case. Lifecycle GHG emissions for the fuel are 80% reduced from the petroleum baseline. Sensitivity analysis shows that reducing MFSP to levels that are competitive with petroleum ($2–3/gal) is challenging, even when considering optimal electricity and CO_2 costs, and carbon credits for this fuel pathway. Utilization of a more efficient route from syngas to ethanol, such as syngas fermentation, which can be in the 90% range, or use of a more selective catalyst for ethanol production, could potentially help reduce costs. Finally, production of high-value synchems could provide more economic impetus than synfuels for integration with LWR operations.

A recent report (Knighton 2020b) studied the framework of options that can incentivize nuclear power plant operations by providing credits for the low-carbon grid power and non-electric products that may be produced from nuclear power, similar to REC. Some conclusions from this report are highlighted below as they apply to synthetic fuel produced by coupling with low-carbon nuclear energy.

Other reports completed by the DOE LWRS program have highlighted the vast and diverse markets for non-electric products that can be produced using nuclear energy (Knighton 2020a, Hu 2019, Frick 2019). The current report has supplemented these studies by providing a first look into two possible pathways for producing synthetic transportation fuels by coupling with low-carbon nuclear energy.

Retiring nuclear plants and not valuing this low-carbon energy with the commensurate credits given to renewable energy may lead to drastic increases in carbon emissions (from substitute baseload plants such as NGCC) at a time when decreases in carbon emissions are being sought, which would be contrary to the goals of decarbonization.

Important points to consider related to possible future low-carbon/zero emissions credit legislation:

- The retention of nuclear power generation is critical to achieving federal and states' decarbonization goals across multiple energy sectors

- Producing hydrogen and other products such as synthetic fuels from nuclear power, especially at low demand periods, increases the capacity utilization factor of NPPs, which can improve the economics of their operation

- The contribution of nuclear power to zero-carbon power markets can be extended further to serve other energy sectors such as transportation, as well as building and industrial heat demand, thus contributing to the goals of decarbonization across multiple energy sectors.

9.   REFERENCES

Alper, E., and Yuksel Orhan, O., 2017, "CO$_2$ utilization: Developments in conversion processes," *Petroleum,* Vol. 3, No. 1, pp. 109–126. doi:https://doi.org/10.1016/j.petlm.2016.11.003

Anghilante, R., Colomar, D., Brisse, A., and Marrony, M., 2018, "Bottom-up cost evaluation of SOEC systems in the range of 10-100 MW," *International Journal of Hydrogen Energy,* Vol. 43, No. 45, pp. 20309–20322, doi:10.1016/j.ijhydene.2018.08.161.

Barger, P., 2002, "Methanol to Olefins (Mto) and Beyond," In *Zeolites for Cleaner Technologies,* Vol. 3, pp. 239–260, Published By Imperial College Press And Distributed By World Scientific Publishing Co.

Becker, W.L., R.J. Braun, M. Penev, and M. Melaina, 2012, ". Production of Fischer-Tropsch liquid fuels from high temperature solid oxide co-electrolysis units," Energy, Vol. 47, pp.: 99–115.

Bipin V. Vorq, E., T.L. Marker, H.R. Nilsen, 1998, "Process for Producing Light Olefins from Crude Methanol," US Patent 5,714,662.

Boardman R.D. et al. 2019. Evaluation of Non-electric Market Options for a Light-water Reactor in the Midwest. INL/EXT-19-55090. Idaho National Laboratory. Idaho Falls, ID

Buttler, A., Koltun, R., Wolf, R., and Spliethoff, H., 2015, "A detailed techno-economic analysis of heat integration in high temperature electrolysis for efficient hydrogen production," *International Journal of Hydrogen Energy,* Vol. 40, No. 1, pp. 38–50. doi:10.1016/j.ijhydene.2014.10.048

Dagle, R.A. and V.M. Dagle. 2020a, "Single-reactor conversion of ethanol to 1-/2-butenes," US Patent 10,647,622.

Dagle, R.A., A.D. Winkelman, K.K. Ramasamy V.L. Dagle, and R.S. Weber, 2020, ". Ethanol as a Renewable Building Block for Fuels and Chemicals," I&EC research, https://dx.doi.org/10.1021/acs.iecr.9b05729.

Dieterich, V., et al. 2020, ". Power-to-liquid via synthesis of methanol, DME or Fischer–Tropsch-fuels: a review," Energy Environ. Sci., DOI: 10.1039/d0ee01187h.

Dutta A., A.H. Sahir, E. Tan, D. Humbird, L.J. Snowden-Swan, P.A. Meyer, and J. Ross, et al. 2015, ". Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbon Fuels: Thermochemical Research Pathways with In Situ and Ex Situ Upgrading of Fast Pyrolysis Vapors," PNNL-23823, Richland, Washington,WA Pacific Northwest National Laboratory.

Dutta, A., M. Talmadge, J. Hensley, M. Worley, D. Dudgeon, D. Barton, P. Groenendijk, D. Ferrari, B. Stears, E.M. Searcy, C.T. Wright, and J.R. Hess, 2011, "Process Design and Economics for Conversion of Lignocellulosic Biomass to Ethanol: Thermochemical Pathway by Indirect Gasification and Mixed Alcohol Synthesis,"". NREL/TP-5100-51400.

Energy Information Agency, 2020a, "Diesel Usage,". diesel usage https://www.eia.gov/energyexplained/diesel-fuel/use-of-diesel.php.

Energy Information Agency. 2020b. Spot prices: https://www.eia.gov/dnav/pet/pet_pri_spt_s1_d.htm; Retail prices: https://www.eia.gov/petroleum/gasdiesel/. Accessed 9/8/2020.

Eng, C. N. A., E.C.; Vora, B.V.. (1998). *Integration of the UOP/HYDRO MTO process into ethylene plants*. Paper presented at the AIChE Spring National Meeting, New Orleans, LA, USA.

Environmental Protection Agency (EPA),). 2020, ". RIN Trades and Price Information," https://www.epa.gov/fuels-registration-reporting-and-compliance-help/rin-trades-and-price-information,. Accessed September 8, /2020.

Frick, K. et al., "Evaluation of Hydrogen Production Feasibility for a Light Water Reactor in the Gasoline Uusage," https://www.eia.gov/tools/faqs/faq.php?id=23&t=10.

Gelbein, A. 2003, PEP Review 2001-11 UOP Methanol to Olefins, SRI Consulting, Menlo Park, California.CA

Giglio, E., Lanzini, A., Santarelli, M., and Leone, P. (2015), "Synthetic natural gas via integrated high-temperature electrolysis and methanation: Part II—Economic analysis," *Journal of Energy Storage,* Vol. 2, pp. 64–79. doi:https://doi.org/10.1016/j.est.2015.06.004

Gogate, M. RMR. 2019, ". Methanol-to-olefins process technology: current status and future prospects," Petroleum Science and Technology. Vol. 37, No. 5, pp. 559–565.

GREET, The Greenhouse Gases, Regulated Emissions, and Energy Use in Transportation, Model 2019.

Griggs, T. (2017), "Low oil prices claim Sasol's proposed $15 billion gas-to-liquids plant near Lake Charles," The Advocate. https://www.theadvocate.com/baton_rouge/news/business/article_3c940178-d051-11e7-8493-07a7ef24d003.html, accessed September 28, 2020.

Hartley, D.S., D.N. Thompson, and H. Cai. 2020, ". Woody Feedstocks 2019 State of Technology Report," INL/EXT-20-57181, Idaho National Laboratory, Idaho Falls, Idaho.ID

Hu, Hongqiang et al., "Technoeconomic Analysis on an Electrochemical Nonoxidative Deprotonation Process for Ethylene Production from Ethane," (December 2019), Idaho National Laboratory, INL/EXT-19-56936,
https://lwrs.inl.gov/Flexible%20Plant%20Operation%20and%20Generation/Technoeconomic_Analysis_on_an_Electrochemical_Nonoxidative_Deprotonation_Process.pdf.

James, R., A. Zoelle, D. Keairns, M. Turner, M. Woods, and N. Kuehn, 2019, "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous coal and Natural Gas to Electricity," NETL-PUB-22638, National Energy Technology Laboratory, Pittsburg, PittsburghPA.

Jasper, S., and El-Halwagi, M. M., 2016, "A Techno-Economic Comparison between Two Methanol-to-Propylene Processes," Vol. 3, No. 684, 2015, *Processes,* Vol. 4, No. 2, doi:ARTN 11 10.3390/pr4020011.

Jones, S. and Y. Zhu, 2009, Techno-economic Analysis for the Conversion of Lignocellulosic Biomass to Gasoline via the Methanol-to-Gasoline (MTG) Process, PNNL-18481, Pacific Northwest National Laboratory, Richland, Washington.

Jones, S., P. Meyer, L. Snowden-Swan, A. Padmaperum, E. Tan, A. Dutta, J. Jacobson, and K. Cafferty. 2013. Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbon Fuels Fast Pyrolysis and Hydrotreating Bio-oil Pathway, PNNL-23053, Pacific Northwest National Laboratory, Richland, WashingtonWA.

Knighton L.T. et al. 2020a. *Scale and Regionality of Nonelectric Markets for U.S. Nuclear Light Water Reactors*. INL/EXT-20-57885. Idaho National Laboratory Idaho Falls, ID

Knighton L.T. et al. 2020b. *Clean Energy Credits for the Production of Low Carbon Hydrogen, Steel and Ammonia using Nuclear Energy*,. INL/EXT-20-58508. Idaho National Laboratory. Idaho Falls, Idaho.ID

Koempel, H., and Liebner, W. (2007). Lurgi's Methanol to Propylene (MTP®) Report on a successful commercialisation. In F. Bellot Noronha, M. Schmal, and E. Falabella Sousa-Aguiar (Eds.), *Studies in Surface Science and Catalysis, Elsevier*, Vol. 167, pp. 261–267.

Lilga, M.; Hallen, R.; Albrecht, K.; Cooper, A.; Frye, J.; Ramasamy, K., " Systems and Processes for Conversion of Ethylene Feedstocks to Hydrocarbon Fuels," U.S.. US 0194257 A1, July 7, 2016.

Lilga, M.A., et al. 2016,. "Systems and processes for conversion of ethylene feedstocks to hydrocarbon fuels," U.S.." US Patent US 2016/0194572 A1 Midwest. (September 2019),). Idaho National Laboratory,

INL/EXT-19-55395, OSTI1569271. DOI: 10.2172/1569271, National Energy Technology Laboratory, 2020. 10.2 Fischer-Tropsch Synthesis. Accessed at https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/ftsynthesis

Lu, Z., Yin, H., Wang, A., Hu, J., Xue, W., Yin, H., and Liu, S., 2016, Hydrogenation of ethyl acetate to ethanol over Cu/ZnO/MOx (MOx=SiO2, Al$_2$O$_3$, and ZrO2) catalysts, *Journal of Industrial and Engineering Chemistry,* Vol. 37, 208–215, doi:https://doi.org/10.1016/j.jiec.2016.03.028.

Lyons, T. W., Guironnet, D., Findlater, M., and Brookhart, M. (2012). Synthesis of p-Xylene from Ethylene. *Journal of the American Chemical Society,* Vol. 134, No. 38, pp. 15708–15711. doi:10.1021/ja307612b

O'Brien, J. E., McKellar, M. G., Stoots, C. M., Herring, J. S., and Hawkes, G. L. (2009), "Parametric study of large-scale production of syngas via high-temperature co-electrolysis," *International Journal of Hydrogen Energy,* 34(9), pp. 4216–4226, doi:10.1016/j.ijhydene.2008.12.021.

Phillips et. al, 2010, Gasoline from Wood via Integrated Gasification, Synthesis, and Methanol-to-Gasoline Technologies, NREL/TP-5100-47594, National Renewable Energy Laboratory, Golden, ColoradoCO.

Phillips, S. et al. 2007, "Thermochemical Ethanol via Indirect Gasification and Mixed Alcohol Synthesis of Lignocellulosic Biomass," NREL/TP-510-41168, National Renewable Energy Laboratory, Golden, Colorado.

Phillips, S. et al. 2007. Thermochemical Ethanol via Indirect Gasification and Mixed Alcohol Synthesis of Lignocellulosic Biomass. NREL/TP-510-41168. National Renewable Energy Laboratory, Golden, CO.

Plasseraud, L., 2010, "Carbon Dioxide as Chemical Feedstock," Edited by Michele Aresta. *ChemSusChem,* Vol. 3, No. 5, pp. 631–632. doi:10.1002/cssc.201000097

Redissi, Y., and Bouallou, C., 2013, "Valorization of Carbon Dioxide by Co-Electrolysis of CO$_2$/H$_2$O at High Temperature for Syngas Production," *Energy Procedia,* Vol. 37, pp. 6667–6678. doi:10.1016/j.egypro.2013.06.599.

Rumayor, M., Dominguez-Ramos, A., and Irabien, A., 2018, "Formic Acid Manufacture: Carbon Dioxide Utilization Alternatives," *Applied Sciences*, Vol. 8, No. 6, doi:10.3390/app8060914.

Process for Ethylene Production from Ethane" (December 2019). Idaho National Laboratory, INL/EXT-19-56936, https://lwrs.inl.gov/Flexible%20Plant%20Operation%20and%20Generation/Technoeconomic_Analysis_on_an_Electrochemical_Nonoxidative_Deprotonation_Process.pdf

Qingxi Fu,* Corentin Mabilat, Mohsine Zahid, Annabelle Brisse and Ludmila Gautier, 2010, ". Syngas production via high-temperature steam/CO$_2$ co-electrolysis: an economic assessment," Energy Environ. Sci., 2010, Vol. 3, pp. 1382–1397. DOI: 10.1039/c0ee00092b.

Samavati, Mahrokh, Andrew Martin, Massimo Santarelli, and Vera Nemanova, 2018, "Synthetic Diesel Production as a Form of Renewable Energy Storage," Energies, Vol. 11, No. 5, https://doi.org/10.3390/en11051223.

Sesto, B., 2016, "Formic Acid," In *Chemical Economics Handbook*: HIS Chemical.

Stevens, R.R., M.M Conway, 1989, "Mixed Alcohols Production from Syngas," US Patent 4,831,060.

Stoots, C., 2010, "Production of Synthesis Gas by High-Temperature Electrolysis of H$_2$O and CO$_2$ (Coelectrolysis),")." Presentation given at the Sustainable Fuels from CO$_2$, H$_2$O, and Carbon-Free Energy conference, Columbia University, New York, New York,NY Accessed at http://energy.columbia.edu/files/2012/11/Coelectrolysis-Rev-2.pdf 9/9/2020.

Summers, Matthew, Matthew Hoffman, Chang-hsien Liao, Matthew Hart, Richard Seiser, Robert Cattolica, Reinhard Rauch, and Matthias Binder, West Biofuels, 2019, ". West Biofuels Fuel Ethanol and Value-Added Chemicals from Biomass Residues," California Energy Commission. Publication Number: CEC-600-2019-140.

Sustainabile Energy Solutions (SES). https://sesinnovation.com/faq.html, accessed September 9, /2020.Yap and J. Lukas, 2015, "Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons via Indirect Liquefaction Thermochemical Research Pathway to High-Octane Gasoline Blendstock Through Methanol/Dimethyl Ether Intermediates," NREL/TP-5100-62402, National Renewable Energy Laboratory, Golden, Colorado.

Tan,, "Conceptual process design and economics for the production of high-octane gasoline blendstock via indirect liquefaction of biomass through methanol/dimethyl ether intermediates," *Biofuels, Bioproducts and Biorefining,* Vol. 10, No.Tan E., L.J. Snowden-Swan, M. Talmadge, A. Dutta, S.B. Jones, K. Kallupalayam Ramasamy, and M.J. Gray, et al. 2016b, ". Comparative Techno-economic Analysis and Process Design for Indirect Liquefaction Pathways to Distillate-range Fuels via Biomass-derived Oxygenated Intermediates Upgrading," Biofuels, Bioproducts *and* Biorefining, Vol. 11, No. 1, pp. :41–66. doi:10.1002/bbb.1710

Tan E., M. Talmadge, A. Dutta, J. Hensley, L.J. Snowden-Swan, D. Humbird, and J. Schaidle, et al. 2016. "Conceptual process design and economics for the production of high-octane gasoline blendstock via indirect liquefaction of biomass through methanol/dimethyl ether intermediates." Biofuels, Bioproducts & Biorefining 10, no. 1:17-35. PNNL-SA-113776. doi:10.1002/bbb.1611

Tan, E., M. Talmadge, A. Dutta, J. Hensley, J. Schaidle, M. Biddy, D. Humbird, L. Snowden-Swan, J. Ross, D. Sexton, R. Yap, and J. Lukas. 2015. Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons via Indirect Liquefaction Thermochemical Research Pathway to High-Octane Gasoline Blendstock Through Methanol/Dimethyl Ether Intermediates. NREL/TP-5100-62402. National Renewable Energy Laboratory, Golden, CO.

Vora, B. V.BV and T.L.TL Marker, ". Process for Producing Light Olefins from Crude Methanol," US 5714662. February 3, 1998.

Weimar, M. R., Chick, L. A., Gotthold, D. W., and Whyatt, G. A. (2013). *Cost Study for Manufacturing of Solid Oxide Fuel Cell Power Systems,* Retrieved from United States: https://www.osti.gov/biblio/1126362.

Xuping Li,*,†,‖ Paul Anderson,‡ Huei-Ru Molly Jhong,‡ Mark Paster,‖ James F. Stubbins,§,‖ and Paul J. A. Kenis. 2016. "Greenhouse Gas Emissions, Energy Efficiency, and Cost of Synthetic Fuel Production Using Electrochemical CO_2 Conversion and the Fischer−Tropsch Process," *Energy Fuels*, 2016, Vol. 30, pp. 5980−5989. DOI: 10.1021/acs.energyfuels.6b00665

Yue, H. R., Ma, X. B., and Gong, J. L. (2014), "An Alternative Synthetic Approach for Efficient Catalytic Conversion of Syngas to Ethanol," *Accounts of Chemical Research,* Vol. 47, 5, pp. 1483–1492, Retrieved from <Go to ISI>://WOS:000336415700004

Zhang, J., Qian, W., Kong, C., and Wei, F., 2015, "Increasing para-Xylene Selectivity in Making Aromatics from Methanol with a Surface-Modified Zn/P/ZSM-5 Catalyst," *ACS Catalysis,* Vol. 5, No. 5, pp. 2982–2988, doi:10.1021/acscatal.5b00192.

Zhu Y., S.A. Tjokro Rahardjo, C. Valkenburg, L.J. Snowden-Swan, S.B. Jones, and M.A. Machinal. 2011, ". Techno-economic Analysis for the Thermochemical Conversion of Biomass to Liquid Fuels," PNNL-19009, Richland, Washington,WA: Pacific Northwest National Laboratory.

Zhou, W., Kang, J. C., Cheng, K., He, S., Shi, J. Q., Zhou, C.,... Wang, Y. (2018,). "Direct Conversion of Syngas into Methyl Acetate, Ethanol, and Ethylene by Relay Catalysis via the Intermediate Dimethyl

Ether,". *Angewandte Chemie-International Edition,* Vol. 57, No. (37), pp. 12012-–12016. doi:10.1002/anie.201807113.

Alper, E., & Yuksel Orhan, O. (2017). CO_2 utilization: Developments in conversion processes. *Petroleum, 3*(1), 109-126. doi:https://doi.org/10.1016/j.petlm.2016.11.003

Anghilante, R., Colomar, D., Brisse, A., & Marrony, M. (2018). Bottom-up cost evaluation of SOEC systems in the range of 10-100 MW. *International Journal of Hydrogen Energy, 43*(45), 20309-20322. doi:10.1016/j.ijhydene.2018.08.161

Barger, P. (2002). METHANOL TO OLEFINS (MTO) AND BEYOND. In *Zeolites for Cleaner Technologies* (Vol. Volume 3, pp. 239-260): PUBLISHED BY IMPERIAL COLLEGE PRESS AND DISTRIBUTED BY WORLD SCIENTIFIC PUBLISHING CO.

Buttler, A., Koltun, R., Wolf, R., & Spliethoff, H. (2015). A detailed techno-economic analysis of heat integration in high temperature electrolysis for efficient hydrogen production. *International Journal of Hydrogen Energy, 40*(1), 38-50. doi:10.1016/j.ijhydene.2014.10.048

Eng, C. N. A., E.C.; Vora, B.V.. (1998). *Integration of the UOP/HYDRO MTO process into ethylene plants*. Paper presented at the AIChE Spring National Meeting, New Orleans, LA, USA.

Giglio, E., Lanzini, A., Santarelli, M., & Leone, P. (2015). Synthetic natural gas via integrated high-temperature electrolysis and methanation: Part II—Economic analysis. *Journal of Energy Storage, 2*, 64-79. doi:https://doi.org/10.1016/j.est.2015.06.004

Jasper, S., & El-Halwagi, M. M. (2016). A Techno-Economic Comparison between Two Methanol-to-Propylene Processes (vol 3, 684, 2015). *Processes, 4*(2). doi:ARTN 11

10.3390/pr4020011

Koempel, H., & Liebner, W. (2007). Lurgi's Methanol To Propylene (MTP®) Report on a successful commercialisation. In F. Bellot Noronha, M. Schmal, & E. Falabella Sousa-Aguiar (Eds.), *Studies in Surface Science and Catalysis* (Vol. 167, pp. 261-267): Elsevier.

Lu, Z., Yin, H., Wang, A., Hu, J., Xue, W., Yin, H., & Liu, S. (2016). Hydrogenation of ethyl acetate to ethanol over Cu/ZnO/MOx (MOx=SiO2, Al_2O_3, and ZrO2) catalysts. *Journal of Industrial and Engineering Chemistry, 37*, 208-215. doi:https://doi.org/10.1016/j.jiec.2016.03.028

Lyons, T. W., Guironnet, D., Findlater, M., & Brookhart, M. (2012). Synthesis of p-Xylene from Ethylene. *Journal of the American Chemical Society, 134*(38), 15708-15711. doi:10.1021/ja307612b

O'Brien, J. E., McKellar, M. G., Stoots, C. M., Herring, J. S., & Hawkes, G. L. (2009). Parametric study of large-scale production of syngas via high-temperature co-electrolysis. *International Journal of Hydrogen Energy, 34*(9), 4216-4226. doi:10.1016/j.ijhydene.2008.12.021

Plasseraud, L. (2010). Carbon Dioxide as Chemical Feedstock. Edited by Michele Aresta. *ChemSusChem, 3*(5), 631-632. doi:10.1002/cssc.201000097

Redissi, Y., & Bouallou, C. (2013). Valorization of Carbon Dioxide by Co-Electrolysis of CO_2/H_2O at High Temperature for Syngas Production. *Energy Procedia, 37*, 6667-6678. doi:10.1016/j.egypro.2013.06.599

Rumayor, M., Dominguez-Ramos, A., & Irabien, A. (2018). Formic Acid Manufacture: Carbon Dioxide Utilization Alternatives. *Applied Sciences, 8*(6). doi:10.3390/app8060914

Sesto, B. (2016). Formic Acid. In *Chemical Economics Handbook*: HIS Chemical.

Weimar, M. R., Chick, L. A., Gotthold, D. W., & Whyatt, G. A. (2013). *Cost Study for Manufacturing of Solid Oxide Fuel Cell Power Systems*. Retrieved from United States: https://www.osti.gov/biblio/1126362

https://www.osti.gov/servlets/purl/1126362

Yue, H. R., Ma, X. B., & Gong, J. L. (2014). An Alternative Synthetic Approach for Efficient Catalytic Conversion of Syngas to Ethanol. *Accounts of Chemical Research, 47*(5), 1483-1492. Retrieved from <Go to ISI>://WOS:000336415700004

Zhang, J., Qian, W., Kong, C., & Wei, F. (2015). Increasing para-Xylene Selectivity in Making Aromatics from Methanol with a Surface-Modified Zn/P/ZSM-5 Catalyst. *ACS Catalysis, 5*(5), 2982-2988. doi:10.1021/acscatal.5b00192

Zhou, W., Kang, J. C., Cheng, K., He, S., Shi, J. Q., Zhou, C.,... Wang, Y. (2018). Direct Conversion of Syngas into Methyl Acetate, Ethanol, and Ethylene by Relay Catalysis via the Intermediate Dimethyl Ether. *Angewandte Chemie-International Edition, 57*(37), 12012-12016. doi:10.1002/anie.201807113